

7th Clean Coal Technology Conference

*21st Century Coal Utilization:
Prospects for Economic Viability, Global
Prosperity and a Cleaner Environment*

PROCEEDINGS Volume II Technical Papers



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TABLE OF CONTENTS

Technical Session I Gasification Systems: How Can CCTs Meet the Needs, Part I

WABASH RIVER IN ITS FOURTH YEAR OF COMMERCIAL OPERATION	1
<i>Steven L. Douglas</i> <i>Dynegy Power Corporation</i>	
TAMPA ELECTRIC COMPANY POLK POWER STATION IGCC PROJECT-PROJECT STATUS	15
<i>John E. McDaniel and Charles A. Shelnut</i> <i>Tampa Electric Company</i>	
OPERATIONAL EXPERIENCE WITH THE 100 MW PINON PINE PROJECT: LESSONS LEARNED AND OUTLOOK FOR THE KRW AIR-BLOWN GASIFICATION TECHNOLOGY WITH HOT GAS CLEANUP	28
<i>Jack W. Motter</i> <i>Sierra Pacific Power Company</i>	
COMMERCIAL-SCALE DEMONSTRATION OF THE LIQUID PHASE METHANOL (LPMEOH™) PROCESS: OPERATING EXPERIENCE UPDATE	46
<i>Barry W. Diamond, Edward C. Heydorn, and</i> <i>Peter J. A. Tijm</i> <i>Air Products and Chemicals Inc.</i>	
<i>Barry T. Street</i> <i>Eastman Chemical Company</i>	
<i>Robert M. Kornosky</i> <i>U.S. Department of Energy, Federal Energy Technology Center</i>	

Technical Session II Combustion Systems: How Can CCTs Meet the Needs, Part II

HEALY CLEAN COAL PROJECT: 1998 COMBUSTOR AND SPRAY DRYER ABSORBER CHARACTERIZATION TESTING	63
<i>Ruth Braswell, Andrew Grove, and Timothy Koyama</i> <i>TRW Space and Technology Division</i>	
<i>John Hardgrove</i> <i>Hardgrove Associates Corporation</i>	
<i>Kevin Redinger and Robert Myers</i> <i>Babcock & Wilcox</i>	

STATUS OF THE LAKELAND McINTOSH UNIT 4 ADVANCED CIRCULATING FLUIDIZED BED COMBINED CYCLE DEMONSTRATION PROJECT	84
--	----

David Adams
Foster Wheeler USA Corporation

Alfred Dodd
Lakeland Electric Utilities

Donald Geiling
U.S. Department of Energy, Federal Energy Technology Center

Archie Robertson
Foster Wheeler Development Corporation

LARGE-SCALE CFB COMBUSTION DEMONSTRATION PROJECT	92
--	----

P.T. Nielsen
JEA

Jerry Hebb
U.S. Department of Energy, Federal Energy Technology Center

Scott L. Darling
Foster Wheeler Energy Corporation

Technical Session III **Beyond 2010: Technology Opportunities and R&D Needs**

ADVANCED GAS TURBINES	109
-----------------------------	-----

William H. Day
Gas Turbine Association and Pratt & Whitney

DEVELOPMENT OF CERAMIC ION TRANSPORT MEMBRANES FOR OXYGEN PRODUCTION	118
---	-----

VanEric E. Stein and Robin E. Richards
Air Products and Chemicals, Inc.

CLEAN COAL TECHNOLOGY IN A CARBON CONSTRAINED WORLD	128
---	-----

Dale Simbeck
SFA Pacific, Inc.

AN ADVANCED COAL GASIFICATION POWER PLANT WITH CO ₂ CONTROL CAPABILITY	141
--	-----

Dwain F. Spencer
SIMTECHE

TECHNICAL SESSION I

Gasification Systems: How Can CCTs
Meet the Needs, Part I

WABASH RIVER IN ITS FOURTH YEAR OF COMMERCIAL OPERATION

Seventh Clean Coal Technology Conference

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ABSTRACT

The Wabash River Coal Gasification Repowering Project (WRCGRP), near Terre Haute, Indiana is now in its fourth year of commercial operation. The 262 MWe project is a joint venture of Dynergy Inc. (formerly NGC Corporation which acquired Destec Energy in 1997) and PSI Energy, a part of Cinergy Corp. The facility is a gasification combined cycle repowering of a 1950s vintage unit at a pulverized coal power plant. The gasification facility utilizes local high sulfur coals (up to 5.9% sulfur) and produces syngas for an advanced combustion turbine as well as sulfur and slag by-products. Wabash River is one of the cleanest, if not the cleanest, of all the coal fired power plants in the world, of any technology.

The Wabash River plant achieved milestones of 10,000 hours of operation and of processing over a million tons of coal late in the third quarter of 1998. This paper will discuss current operations and operational challenges as well as briefly recapping the current status of some problematic areas from earlier plant operation.

I. INTRODUCTION

Since its inception, the Wabash River Coal Gasification Repowering Project Joint Venture has maintained a determined focus throughout its continued successful demonstration of the application of an advanced Clean Coal Technology in an established utility setting. Thanks to the foresight of and funding by the United States Department of Energy, the Project participants have continued to overcome technical barriers in the pursuit of the original project objectives, those objectives being,

- for PSI Energy, to establish a clean, low cost, energy efficient baseload capacity addition that would serve as a substantial element in their plan to comply with the Clean Air Act requirements;
- for Dynergy, to continue the development and advancement of its gasification technology to the next generation to enhance the competitive position for future IGCC applications;
- for the U.S. DOE, to demonstrate the commercial scale application of a clean coal technology that would provide utility and industrial decision makers with economic, environmental and social data as to the competitiveness of clean coal technologies to meet their power generation needs for the next millennium.

Through the 262-MWe Wabash River Coal Gasification Repowering Project, the objectives of the participants are being met as availability-limiting factors are mitigated, plant operating and maintenance costs are being reduced and feedstock flexibility is being demonstrated. The Wabash River Project is demonstrating a highly efficient, environmentally superior clean coal technology that has done much to abate the barriers to further commercialization of this technology in a variety of settings. Findings from the nearly four years of operation of the Wabash River facility are also being leveraged into the next generation of power and chemical production megaplexes as Dynegy participates in the U.S. DOE's "Vision 21" program as well as continues to participate in various commercial project proposals that would perhaps enhance the more near-term commercial viability of IGCC power production.

II. OVERVIEW

The Project participants, Dynegy Power Corp. (Dynegy), of Houston, Texas, and PSI Energy, Inc. (PSI), of Plainfield, Indiana, formed the joint venture to participate in DOE's Clean Coal Technology (CCT) program to demonstrate the coal-gasification repowering of an existing generating unit impacted by the Clean Air Act. The participants jointly developed, separately designed, constructed, own, and are now operating an integrated coal-gasification combined-cycle power plant, using Dynegy's coal gasification technology to repower the oldest of the six units at PSI's Wabash River Generating Station in West Terre Haute, Indiana. Dynegy's gasification process is integrated with a new General Electric 7 FA combustion turbine generator and a heat recovery steam generator in the repowering of a 1950s-vintage Westinghouse steam turbine generator using some pre-existing coal handling facilities, interconnections, and other auxiliaries.

The Project is currently in its fourth year of operation and near the end of the planned demonstration period under the DOE CCT program. The Project, which is the world's largest single-train coal-gasification combined-cycle plant operating commercially, has demonstrated through its early operation, the ability to run at full load capability while meeting the environmental requirements for sulfur and NO_x emissions. Cinergy, PSI's parent company, dispatches power from the Project, with a demonstrated heat rate of under 9,000 Btu/kWh (HHV), second only to their hydroelectric facilities on the basis of environmental emissions and efficiency.

III. BACKGROUND

Dynegy Gasification Technology Evolution

The development of the Dynegy gasification process began in the early 1970s. The original technology owner, Dow Chemical, wanted to diversify its fuel base from natural gas to lignite and coal for its power-intensive chlor-alkali processes and began to develop the gasification process through basic R&D and pilot plants. The first commercial gasification plant, Louisiana Gasification Technology, Inc. (LGTI), followed in Plaquemine, LA. This project operated from the second quarter 1987 until the third quarter 1995 under subsidy from the United States Synthetic Fuels Corporation (and later the U. S. Treasury Department). When Destec Energy was formed in 1989, the gasification technology was transferred from Dow Chemical to Destec. In June of 1997, Destec Energy was purchased by and became a wholly owned subsidiary of NGC Corporation, a leading gatherer, processor, transporter and marketer of energy products and services in North America and select markets worldwide. In June of 1998, NGC Corporation changed its name to Dynegy Inc.

Wabash River Project Development

Destec, now Dynegy, approached PSI in early 1990 to initiate discussions concerning the DOE Clean Coal Technology Round IV program solicitation. Through the Wabash River Coal Gasification Repowering Project Joint Venture, the project submittal was made. In September 1991, the Project was among nine projects selected from 33 proposals. The Project was selected to demonstrate the integration of Dynegy's gasification process with a new GE 7 FA combustion turbine generator and a heat recovery steam generator (HRSG) in the repowering of an aged steam turbine generator to achieve improved efficiency and reduced emissions.

Project Organization, Commercial Structure, and Costs

There are two major agreements which establish the basis of the Project. First, the Joint Venture Agreement was created between PSI and Dynegy to form the Wabash River Coal Gasification Repowering Project Joint Venture in order to administer the Project under the DOE Cooperative Agreement. Second, the Gasification Services Agreement (GSA) was developed between PSI and Dynegy and contains the commercial terms under which the Project was developed and is now operated.

PSI Responsibilities:

- build power generation facility to an agreed schedule
- own and operate the power generation facility
- furnish Dynegy with a site, coal, electric power, storm water and wastewater facilities, and other utilities and services.

Dynegy Responsibilities:

- build gasification facility to an agreed schedule
- own and operate the gasification facility
- guarantee operating performance of the coal gasification facility, including product and by-product quality
- deliver syngas and steam to the power generation facility

IV. REVIEW OF TECHNOLOGY

General Design and Process Flow

The Dynegy coal gasification process (Figure 1) features an oxygen-blown, continuous-slugging, two-stage, entrained-flow gasifier which uses natural gas for heat-up. Coal is milled with water in a rod mill to form a slurry. The slurry is combined with oxygen in mixer nozzles and injected into the first stage of the gasifier, which operates at 2600°F and 400 psig. Oxygen of 95% purity is supplied by a turnkey, Air Liquide, 2,060-ton/day low-pressure cryogenic distillation facility which Dynegy owns and operates.

In the first stage, coal slurry undergoes a partial oxidation reaction at temperatures high enough to bring the coal's ash above its melting point. The fluid ash falls through a taphole at the bottom of the first stage into a water quench, forming an inert vitreous slag. The syngas then flows to the

second stage, where additional coal slurry, but no additional oxygen, is injected. This coal reacts endothermically with the hot syngas to enhance syngas heating value and to improve overall efficiency.

The syngas then flows to the high-temperature heat-recovery unit (HTHRU), essentially a firetube steam generator, to produce high-pressure saturated steam. After cooling in the HTHRU, particulates in the syngas are removed in a hot/dry filter and recycled to the gasifier where the carbon in the char is converted into syngas. After particulate removal, the syngas is further cooled in a series of heat exchangers, is water scrubbed for chlorides removal and is passed through a catalyst which hydrolyzes carbonyl sulfide into hydrogen sulfide. Hydrogen sulfide is removed using methyldiethanolamine (MDEA) based absorber/stripper columns. The “sweet” syngas is then moisturized, preheated and piped over to the power block.

The key elements of the power block are the General Electric MS 7001 FA high-temperature combustion turbine (CT) / generator, the heat recovery steam generator (HRSG), and the repowered steam turbine. The GE 7 FA is a dual-fuel turbine (syngas for operations and No. 2 fuel oil for startup) capable of a nominal 192 MW when firing syngas, attributed to the increased mass flows inherent with syngas use. Steam injection is used for NO_x control, but the steam flow requirement is minimal compared to that of conventional systems because the syngas is moisturized at the gasification facility, making use of low-level heat in the process. The water consumed in this process is continuously made up at the power block by water treatment systems which clarify and treat river water.

The HRSG for this project is a single-drum design capable of superheating 754,000 lb/hr of high-pressure steam at 1010°F and 600,820 lb/hr of reheat steam at 1010°F when operating on design-basis syngas. The HRSG configuration was specifically optimized to utilize both the gas-turbine exhaust energy and the heat energy made available in the gasification process. The nature of the gasification process in combination with the need for strict temperature and pressure control of the steam turbine led to a great deal of creative integration between the HRSG and the gasification facility.

The repowered unit, originally installed in 1952, consisted of a conventional coal-fired boiler feeding a Westinghouse reheat steam turbine rated at 99 MW but derated in recent years to 90 MW for environmental reasons. Repowering involved refurbishing the steam turbine to both extend its life and withstand the increased steam flows and pressures associated with the combined-cycle operation. The repowered steam turbine produces 104 MW which combines with the combustion turbine generator’s 192 MW and the system’s auxiliary load of approximately 34 MW to yield 262 MW (net) to the Cinergy grid.

The Air Separation Unit (ASU) provides oxygen and nitrogen for use in the gasification process but is not an integral part of the plant thermal balance. The ASU uses services such as cooling water and steam from the gasification facilities and is operated from the gasification plant control room.

The gasification facility produces two commercial by-products during operation. Sulfur, removed as 99.99 percent pure elemental sulfur, is marketed to sulfur users, shipped in rail tank cars as a liquid. Slag is targeted as an aggregate in asphalt roads and as structural fill in various types of construction applications.

Gasification Process

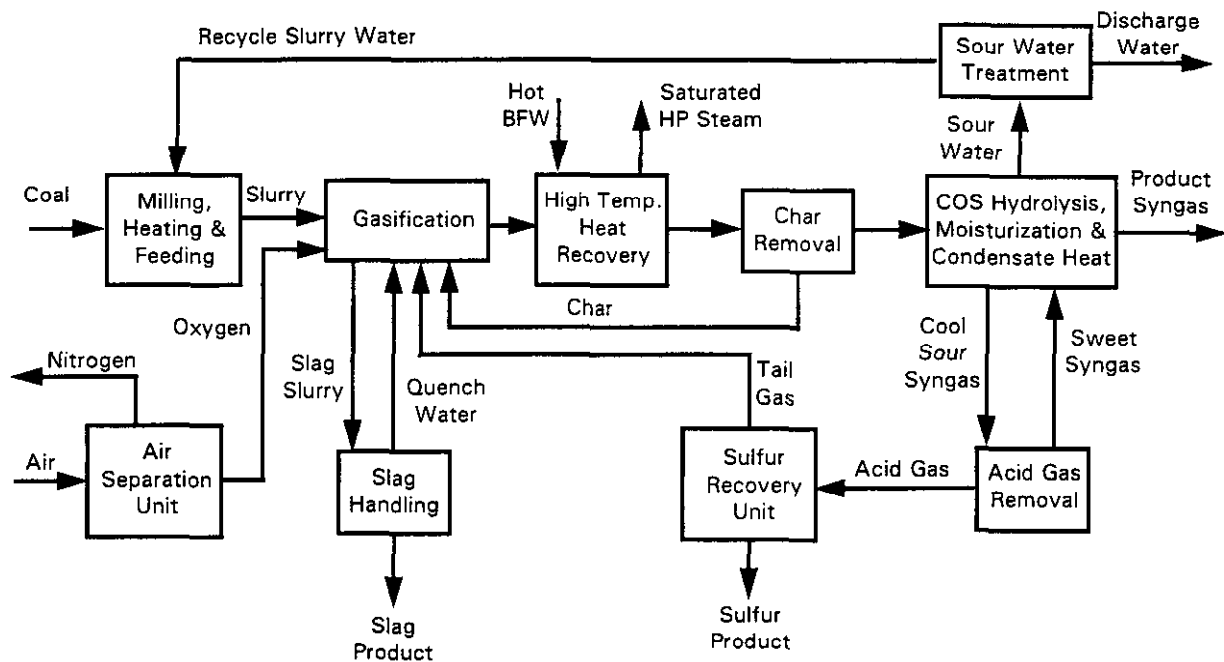


FIGURE 1

Technical Advances

Using integrated coal gasification combined-cycle technology to repower a 1950s-vintage coal-fired power generating unit essentially demonstrates a technical advance in and of itself.

More specifically, high energy efficiency and superior environmental performance while using high sulfur bituminous coal are the result of several improvements to Dynegy's gasification technology, including:

- **Hot/Dry Particulate Removal**, applied here at full commercial scale.
- **Syngas Recycle**, which provides fuel and process flexibility while maintaining high efficiency.
- A **High-Pressure Boiler**, which cools the hot, raw gas by producing steam at a pressure of 1,600 psia.
- A **Dedicated Oxygen Plant**, which produces 95% pure oxygen for use by the Project. Use of 95% purity increases overall efficiency of the Project by lowering the power required for production of oxygen.
- **Integration of the Gasification Facility with the Heat Recovery Steam Generator** to optimize both efficiency and operating costs.
- The **Carbonyl Sulfide Hydrolysis** system, which allows such a high percentage of sulfur removal.
- The **Slag Fines Recycle System**, which recovers carbon remaining in the slag by-product stream and recycles it back for enhanced carbon conversion. This also results in a higher quality by-product slag.

- **Fuel Gas Moisturization**, which uses low-level heat to reduce steam injection required for NO_x control.
- **Sour Water Treatment and Tail Gas Recycle**, which allow more complete processing of combustible elements, thereby reducing waste water discharge and gaseous emissions.

V. OPERATIONS EXPERIENCE

The plant achieved commercial operation status in November of 1995 after a three month start-up phase that included equipment shakedown and performance testing.

The first commercial year of operation of the WRCGRP saw the plant build on the success of the start-up period, with primary focus on attaining maximum sustained capacity for the purpose of final performance testing for the ASU Facility and Gasification Plant.

As the Project accumulated the early run time, evaluation of the technical advances that are a part of this demonstration facility showed that most of the new unit operations performed very well. However, two of the areas contributed problems that significantly affected run time.

The first problem area was the reliability of the particulate removal system, primarily due to breakage of ceramic candle filters in the primary particulate removal vessels. The second problem area was chloride concentrations in both the COS hydrolysis catalyst beds and downstream heat exchangers in the syngas cooler line-up. Unexpected localized high chloride concentrations contributed to catalyst poisoning and chloride stress corrosion cracking in the syngas heat exchangers. Within the gasification plant, a large scale capital improvement project was launched early in the first commercial year to reduce downtime related to these two severe problems as well as address other, less severe process-related problems. An aggressive implementation schedule targeted these improvements for late in the first commercial year in order to maximize impact on second commercial year operating rate. A discussion of these improvements and their positive impact on second and third years' operations follows in the area operations summaries below.

On the power block side, the new advanced combustion turbine (CT) has performed very well on syngas. The turbine's operation has been more stable on syngas than on oil. The blade temperatures have been more evenly distributed and have had less temperature spiking. NO_x is reduced with steam injection and has been adjusted to meet air permit requirements. The turbine experienced three areas of additional work after the acceptance of syngas. The first was in the syngas module and the piping from the module to the gas turbine. Expansion bellows required redesign and replacement to eliminate cracking in the flow sleeves. This problem was corrected by GE efforts in early syngas runs. The second problem is the syngas purge control. These problems were primarily related to field devices such as solenoid valves and flow measuring devices. The solenoids have been redesigned and replaced and GE continues to work on flow measuring devices. The third area was the GE required 2-3 spacer modifications.

The second year of commercial operation identified cracking problems with the combustion turbine combustion liners. Several outages resulted to allow weld repair of cracked liners. The cracking was located near the head end of the liner and around cooling holes. Evaluation of cause resulted in a replacement of the fuel nozzles and liners as a warranty item for GE.

Also, in the second year of operation, tube leaks in the HRSG superheater and reheater area became a degrader of availability. The cause of the tube leaks was determined to be the limiting of needed expansion during startup conditions. A change to the main steam piping support system was made. Also, a change was made in the boiler roof/penthouse floor to allow for better expansion of the roof panels to reduce the stress created on the vertical tubing that results from the binding roof panels.

Recently, in mid-March 1999, damage was incurred to rows 14 through 17 of the air compressor rotor and stator of the combustion turbine assembly which, due to repair parts availability and shop time, has resulted in a lengthy unscheduled outage for the Project. At the time of this writing, PSI Energy, their insurer and GE were in the process of performing root cause failure studies, with no outcome agreed upon at this point.

Operations Statistics

GASIFICATION PLANT PRODUCTION STATISTICS						
Time Period	Longest Run (Days)	On Coal (Hours)	Coal Processed (Tons)	On Spec. Syngas Produced (MMBtu)	Sulfur Produced (Tons)	Equivalent SO ₂ Captured (Mlbs)
Startup through 12/95	3	505	45,166	254,521	560	2,370
1996 Calendar Year	19	1,902	184,380	2,769,683	3,299	13,183
1997 Calendar Year	46	3,885	387,501	6,214,864	8,607	34,392
1998 Calendar Year	82	5,279	561,494	8,831,011	12,452	49,756
1999 through 5/99 ♦	20	851	93,969	1,459,168	2,096	8,375
Overall Totals		12,422	1,272,510	19,529,247	27,014	108,076

♦ Statistics through 05/31/99 Note: Combustion Turbine was unavailable from 3/14/99 through 6/15/99 (estimate).

TABLE I

COMBINED CYCLE PLANT PRODUCTION STATISTICS				
Time Period	Total Combustion Turbine Operation (Hours)	Total Combustion Turbine Operation on Syngas (Hours)	Gross Power Generation from Syngas (MWH)	Highest Capacity Operation (MWH, Gross)
1996 Calendar Year	2,177	1,553	278,164	296
1997 Calendar Year	4,261	3,701	940,365	296
1998 Calendar Year	5,763	5,138	1,429,155	296
1999 through 5/99 ♦	986	820	229,814	296
Overall Totals	13,187	11,212	2,877,498	

♦ Statistics through 05/31/99 Note: Combustion Turbine was unavailable from 3/14/99 through 6/15/99 (estimate).

TABLE II

During the third year of commercial operation Dynegy demonstrated operation on a second coal feedstock as well as a blend of two different Illinois No. 6 coals. This ability to blend coal feedstocks has improved the fuel flexibility for the site.

In addition to the coal experience, the Wabash plant conducted a very successful petroleum coke (petcoke) test run at the end of the second operating year. The petroleum coke test was performed to demonstrate the fuel flexibility of Dynegy's Gasification Technology. During the test, over 18,000 tons of high sulfur, delayed petroleum coke were gasified to produce 350,000 MMBtu of synthetic gas that was fed to the combustion turbine. No process modifications were made to accommodate the change in feedstock and no negative effect was realized from processing the petroleum coke.

Overall thermal performance (see Table III) was slightly improved during petcoke operation, with overall plant efficiency at 40.2% (HHV). The reduced syngas consumption by the combustion turbine in the "actual" cases reflects the actual turbine performance, which was better than the manufacturer's guarantee.

In the "actual" cases in Table III, overall plant electrical production has been adjusted to compensate for the steam turbine output lost due to the gas turbine HRSG feedwater heater which has been out of service. This results in the HRSG failing to capture as much heat from the turbine exhaust as expected, as well as requiring additional steam for deaeration, which reduces the steam turbine output.

THERMAL PERFORMANCE SUMMARY			
	Design	Actual	
	Coal	Coal	Petcoke
Nominal Throughput, tons/day	2550	2450	2000
Syngas Capacity, MMBtu/hr	1780	1690	1690
Combustion Turbine MW	192	192	192
Steam Turbine MW	105	96	96
Auxiliary Power MW	35	36	36
Net Generation, MW	262	261	261
Plant Efficiency, % (HHV)	37.8	39.7	40.2
Sulfur Removal Efficiency, %	>98	>99	>99

TABLE III

The following is an operations summary of each major operating area, including the areas mentioned above, with a discussion of process modifications incorporated to date.

Area Operations Summaries

Coal Slurry Preparation, Storage and Feed. Coal is ground into a slurry in a rod mill, using recycled water from the gasification process. Wet milling reduces fugitive particulate emissions and minimizes water consumption and effluent waste water volume. The slurry is stored in an agitated tank large enough to supply the gasifier needs during forced rod mill outages.

The Slurry Preparation Area has processed over one and one quarter million tons of coal with no significant problems. The Slurry Storage and Feed System has also performed very well since the beginning. In fact, only a few hours of downtime since start-up can be directly attributed to these two systems. Table IV depicts the typical coal processed as well as the petroleum coke processed during the November 1997 test.

FUEL ANALYSIS		
Analysis	Typical Coal	Petroleum Coke
Moisture, %	15.2	7.0
Ash, %	12.0	0.3
Volatile, %	32.8	12.4
Fixed Carbon, %	39.9	80.4
Sulfur, %	1.9	5.2
Heating Value, as Rec'd, Btu/lb	10,536	14,282

TABLE IV

Oxygen/Nitrogen Generation and Supply. The Air Separation Unit, supplied by Liquid Air Engineering Co., produces 2060 tons/day of oxygen at 95% purity as well as high-purity nitrogen and dry process air for use in the gasification process. The process involves air compression, purification, cryogenic distillation, oxygen compression, and a nitrogen storage and handling system. The ASU was started up in April of 1995 and has reliably supplied products to the gasifier island, although instrumentation related nuisance trips continue to occur periodically. A large effort is underway in 1999 to reduce the number of these nuisance trips.

Gasification and Slag Handling. The gasification and slag handling areas have continued to perform well. The slag removal system has continued to operate essentially trouble free. The gasifier has consistently processed the coal into high-quality syngas. The taphole from which the slag by-product is removed from the gasifier has experienced remote instances of plugging, but unrelated to gasifier performance. Although the slag handling system has performed well, improvements have been identified to improve the quality and consistency of the slag by-product, thus improving the marketability. The high pressure slurry burners have required replacement approximately every 40-50 days. However, the facility availability impact has been minimal since burners can be changed in less than 18 hours coal-to-coal. As with other availability-limiting components of the process, an effort is ongoing to improve the run life of the slurry burners.

Syngas Cooling, Particulate Removal, and COS Hydrolysis. During the first commercial operating year, this area of the gasification plant experienced problems which can be summarized into three areas: (1) ash deposition at the inlet to the firetube boiler, (2) particulate breakthrough in the barrier filter system, and (3) poisoning of the COS catalyst due to chlorides and metals in the syngas. It was these problems that necessitated a large-scale capital improvement program initiated early in 1996.

Ash deposition has not been a big contributor to overall downtime, but did limit run time on several occasions into the second commercial year due to deposition at the inlet to the waste heat boiler tubes. A major improvement was implemented in the third quarter of 1997. This improvement modifies hot gas path flow geometry and velocities so as to minimize large-scale deposits which can spall off to produce deposition within the waste heat boiler. Management of the ash that does reach the boiler has been improved such that the boiler now remains clean for extended run lengths.

Particulate breakthrough within the barrier filter system experienced during the first commercial year was primarily due to movement and breakage of the ceramic candle filter elements. Substantial downtime is associated with entry into the particulate filter vessels. Therefore, the improvement projects identified early in 1996 placed significant emphasis on improvements to this system to eliminate particulate breakthrough. These improvements, including replacement of the ceramic elements with metallic candles, were implemented during the fourth quarter of 1996 and have proven successful.

Downtime associated with the barrier filtration system has been reduced by nearly 80% over the first commercial year statistics. The single gasification plant outage during the second commercial year resulting from candle element failure was directly related to a failure within the pulse valve system. Consequently, the barrier filtration system has accounted for less than 16 days of outage time due to candle element problems in 1997 vs. well over 100 days in 1996. Most of the barrier filtration downtime in 1997 was a result of filter element blinding which required off-line cleaning. An ongoing filter element development program targeted at improving filter metallurgy, blinding rates, and cleaning techniques has reduced 1998 downtime associated with the particulate removal system by an additional 66% over 1997. As a result of these continuing efforts, downtime associated with this system decreased in each of the four quarters of 1998.

To further maximize the availability of the particulate removal system and minimize maintenance costs, the plant has installed a slipstream unit capable of testing alternate filter element materials as well as process operating condition effects on element conditioning and overall life. Since commissioning in the fourth quarter of 1997, the unit has logged over 2250 coal operating hours while successfully completing nine test campaigns consisting of seven different filter evaluations for blinding, corrosion and mechanical integrity. The initial slipstream test results were presented at a DOE (FETC) sponsored conference in July, 1998, in a presentation titled "Particulate Filters at the Wabash River Coal Gasification Repowering Project". The DOE continues to be very involved with and supportive of Dynegy's efforts to improve the reliability and operating economics of the barrier filtration system which can potentially have widespread application in IGCC systems.

Poisoning of the COS hydrolysis catalyst due to chlorides and metals led to early replacement of the catalyst. To address this concern as well as metallurgy concerns with chlorides further downstream in the process, a wet chloride scrubber system was installed during September of 1996 as the first phase of process improvements. Since start-up in October of 1996, this system has performed per design in the removal of chlorides from the syngas and has eliminated poisoning concerns within the hydrolysis catalyst as well as corrosion concerns in the downstream equipment. An additional target of the process improvement plan was the identification of an alternate hydrolysis catalyst, less prone to poisoning from both chlorides and trace metals within the syngas. Alternate catalyst was identified and installed in October of 1997 and has proven high performance in the hydrolysis process with minimal degradation in performance over extended run time.

Low Temperature Heat Recovery and Syngas Moisturization. Since the installation of the new chloride scrubbing system late in the first commercial year, this section of the process has performed well in terms of providing the moisturization for the syngas and providing heat transfer as designed.

Acid Gas Removal and Sulfur Recovery. The acid gas removal process has effectively demonstrated removal of over 99% of the sulfur in the syngas, with overall sulfur recovery at better than 98%. The typical sweet syngas composition from the plant has been consistent and is shown in Table IV. The other permitted air emission sources are the combustion turbine exhaust and the syngas startup/shutdown flare.

PRODUCT SYNGAS ANALYSIS		
Analysis	Typical Coal	Petroleum Coke
Nitrogen, Vol. %	1.9	1.9
Argon, Vol. %	0.6	0.6
Carbon Dioxide, Vol. %	15.8	15.4
Carbon Monoxide, Vol. %	45.3	48.6
Hydrogen, Vol. %	34.4	33.2
Methane, Vol. %	1.9	0.5
Total Sulfur, ppm _v	68	69
Higher Heating Value, Btu/SCF	277	268

TABLE IV

Environmental Performance. Total sulfur dioxide (SO₂) emissions from the three permitted emissions points (HRSG stack, gasification flare stack, and tail gas incinerator stack) have demonstrated the ability of the gasification process to successfully operate below 0.1 lbs SO₂ emitted per MMBtu of coal input. To date, emission rates as low as 0.03 lbs/MMBtu have been attained. This represents better than a 94% reduction in SO₂ emissions from the decommissioned Unit 1 boiler at the Wabash River Generating Station. The 0.1 lbs/MMBtu is significantly below acid rain limits set for the year 2000 at 1.2 lbs/MMBtu under the Clean Air Act Amendment. Through May of 1999, the Project has captured approximately 108 million pounds equivalent of sulfur dioxide emissions as 99.99% pure elemental sulfur.

Combustion Turbine. The combustion turbine has operated in excess of 13,000 fired hours on syngas and No. 2 fuel oil. The CT has also occasionally operated in a simple cycle configuration (without the gasification plant) as a liquid fuel-fired combined-cycle peak-service generator.

VI. SUMMARY / RECENT ACTIVITIES

During the third and into the fourth commercial years of operation, the Wabash River Coal Gasification Repowering Project has continued to make progress toward achieving the project goals. Both the gasification and combined-cycle plants have demonstrated the ability to run at capacity and well within environmental compliance while using locally mined high sulfur Illinois Basin bituminous coals. The experience with bituminous coals and petroleum coke at Wabash River has

added to the fuel flexibility portfolio of Dynegy, which had previously processed both lignite (from Texas and Louisiana) and subbituminous coals (from Montana, Utah, and Wyoming) during its earlier process development efforts at other facilities.

Early identification of availability limiting process problems within the gasification plant led to aggressive implementation of improvement projects which resulted in 224% more syngas produced during the second year than in year one. The syngas produced during the third year exceeded the second year's production by an additional 42%. Also, the plant achieved 77% availability for the third commercial year (omitting downtime attributed to combined cycle power generation and alternative fuel testing). Further analysis of downtime contributors and subsequent modifications, as well as indicated barrier filter slip stream testing will further improve plant availability and reduce operating and maintenance (O&M) costs.

Additionally, as plant performance continued to improve on all levels throughout 1998 and into 1999, several project milestones were reached in what are termed the plant's "Big Four Goals" of Safety, Environmental, Production, and O&M Spending:

Began running new coal feed (Miller Creek)	June 1998
Completed 14 Months OSHA Recordable-free	September 1998
Surpassed 1,000,000 tons of coal processed	September 1998
Surpassed 10,000 hours of coal operation	September 1998
Began demonstrating operation with blended feeds	September 1998
Back-to-back trillion Btu months (<i>Second time in 1998!</i>)	November and December 1998
Reduced routine O&M spending >25%	From 1996 to 1998
Record <i>Commercial</i> Operating Year (12/1/97-11/30/98)	1998 (9.13 trillion Btu produced)
Surpassed 100,000,000 pounds equivalent of SO ₂ captured	January 1999

While the beginning of 1999 had all the early indications of another record syngas production year, the combustion turbine compressor incident in March 1999 has brought about a different course of activities in the Coal Gasification and Air Separation facilities. The unplanned three-month outage has provided Dynegy personnel with the opportunity to accelerate the installation of many additional minor plant improvement projects and safety enhancements that would have otherwise been scheduled for a plant outage in the fall of 1999. Also, many routine and preventive maintenance tasks to bring equipment back to prime operating condition were carried out during this period. The Air Separation Unit was taken through its first major derime (heat up to greater than ambient conditions to purge carbon dioxide and trace hydrocarbons) since its startup in early 1995. Many, many "back burner" items were also addressed. To keep O&M costs in check through improved overall maintenance productivity, the normal operations and maintenance personnel work schedules were shifted to four 10-hour days during the unplanned downtime. Much of the maintenance work was accomplished by operations personnel, supplementing the contract craftsman workforce.

The production lull also afforded Dynegy personnel many training opportunities, for both new and refresher courses, allowed additional control system software enhancements, provided an opportune window for the implementation of both a new computerized maintenance management system and Y2K solutions and allowed for some strengthening of the teamwork among the various functions at

Wabash River through some planned offsite “team building” activities. When the facility is brought back on line in June 1999, it is anticipated that it will be the beginning of a new record production run that will extend through the peak electrical demand months of the summer.

VII. OUTLOOK

The current trend that is most significantly impacting the advancement of gasification technology in the global market is the incentive to utilize “low value” feeds such as petroleum coke and residual oil. The current economics of coal based gasification for power do not compare favorably to natural gas generation in areas where natural gas is available and is at relatively low costs. This has prompted Dynegy to pursue alternate feedstocks and to perform the petroleum coke testing at Wabash River, discussed in this paper, and documented further in the paper entitled “Alternate Fuel Testing at the Wabash River Coal Gasification Repowering Project”.

The conclusion from the petroleum coke testing at Wabash River is that operation was not significantly different than coal operation, and that the equipment and systems put in place at Wabash River were adequate for this operation without modification. Gasifier operation was extremely successful, even with flux rates as low as 2%, and the petroleum coke trace metal constituents were effectively captured in the slag produced. There was a negligible amount of tar formation. No problems were encountered in the operation of the dry char particulate removal system, despite a higher dust loading. Thermal efficiencies greater than 40% HHV have been achieved with “F” class combustion turbines and future facilities should be able to approach 50% efficiency with the advanced “H” class turbines.

It appears that future units designed to utilize petroleum coke as their primary fuel source can be implemented with similar systems as Wabash River, but with some improvements to reduce capital costs or improve operability. Low flux requirements demonstrated at Wabash River mean that the slag, ash and flux systems in future plants can be downsized considerably. The low reactivity of the petroleum coke will mean elimination of certain equipment installed at Wabash River intended to minimize tar formation. The coal handling and slurry preparation systems at petcoke-fired facilities can be downsized as well. Operation should continue to be as smooth as with coal.

Another trend in the market is the increasing use of syngas for the manufacture of value-added chemicals or transportation fuels. Several well-proven gasification plants are currently operating in the world today producing various chemical products from a wide variety of feedstocks. Others are coming on line in the new few years. (Reference: www.gasification.org). Due to the current natural gas market conditions, it very well could be that the near-term niche for gasification lies not so much in the production of electricity, but possibly in a tailored tri-generation operation, where electricity, steam and chemicals are economically “bundled” as products from a fully integrated complex.

In late 1998, Cinergy Corp., through its operating company PSI Energy, Inc., reached agreement to purchase the Gasification Services Contract with Dynegy, subject to regulatory approval. This agreement allows Cinergy to purchase the remaining term of the 25-year contract, which has become “out-of-market” in comparison to today’s natural gas fuel market. Dynegy, in conjunction with Cinergy and the Department of Energy, are exploring alternatives for continued operation of Wabash River in a more “market-based” mode, and potentially with a new technology owner. With

this development and the ongoing efforts to improve the commercial viability of the Wabash River Project, the focus has sharpened on how to make the technology competitive in today's market.

Building on the lessons learned and the many successes to date, the Wabash River Coal Gasification Repowering Project looks forward to continued demonstration of the viability of the technology in an expanding variety of applications. Consideration is being given for additional alternate fuel tests in conjunction with continued emphasis in improvement of operating rate and lowering of syngas production costs. Primary emphasis will be placed on the competitive cost of syngas and power from Wabash River as compared to alternatives available in the current and future markets. With the early major plant problems now resolved and additional process enhancements in place, the focus of the personnel at Wabash River remains steadfast - to lower O&M costs while making incremental improvements in safety performance, plant availability and plant capacity factor. As a result of this ongoing focus, the technology demonstrated at the Wabash River facility is well positioned to provide the solution to the growing global demand for efficient, environmentally superior, competitive energy conversion to power from solid carbon-based feedstocks. Additionally, efforts will be expended by Dynegy personnel to incorporate other imaginative technologies and to pursue value-added uses for syngas produced from coal (and/or other feeds) such as is envisioned through forward-thinking concepts like the DOE's "Vision 21" initiative. In the face of the current power market challenges brought about by abundant and low cost natural gas, Wabash River personnel will aggressively use their collective ingenuity, addressing opportunities on a variety of fronts, to propel the Dynegy Gasification Technology to the forefront as an economically viable coal-based alternative for electrical power production - and a whole lot more!

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TAMPA ELECTRIC COMPANY POLK POWER STATION IGCC PROJECT--PROJECT STATUS

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ABSTRACT

Over the last nine years, Tampa Electric Company has taken the Polk Power Station from a concept to a reality. In 1996, we reported on the permitting, engineering, construction, contracting, and staffing of the project. Our papers at the Fifth and Sixth Annual Clean Coal Technology Conferences discussed startup experiences, the early operating history of the plant, and results of alternate fuel testing. In this year's paper, we will provide an update on the reliability statistics and discuss in more detail some of the specific problems we have encountered and resolved.

I. BACKGROUND

Participants

Tampa Electric Company (TEC) is the owner and operator of Polk Power Station. TEC is an investor-owned electric utility, headquartered in Tampa, Florida. TEC has about 3650 MW of generating capacity. Over 97 percent of TEC's power is produced from coal. TEC serves over 500,000 customers in an area of about 2,000 square miles in west-central Florida, primarily in and around Tampa, Florida. TEC is the principal, wholly-owned subsidiary of TECO Energy, Inc., an energy related holding company.

TECO Power Services (TPS), another subsidiary of TECO Energy, Inc., provided project management services for Polk Power Station during its design, construction, and startup phases. TPS is now concentrating on commercialization of this IGCC technology as part of the Cooperative Agreement with the U.S. Department of Energy. TPS was formed in the late 1980's to take advantage of the opportunities in the non-regulated utility generation market. TPS currently owns and operates 2 natural gas fired power plants, a 295 MW plant in Hardee County, Florida, and a 78 MW plant in Guatemala. In addition, TPS has several other projects at various stages of development.

The project is partially funded by the U.S. Department of Energy (DOE) under Round III of its Clean Coal Technology Program. The research was sponsored by the U.S. Department of Energy's Federal Energy Technology Center, under contract DE-FC-21-91MC27363 with Tampa Electric Company, PO Box 111, Tampa, FL 33601; Fax: 941-428-5927

Objectives

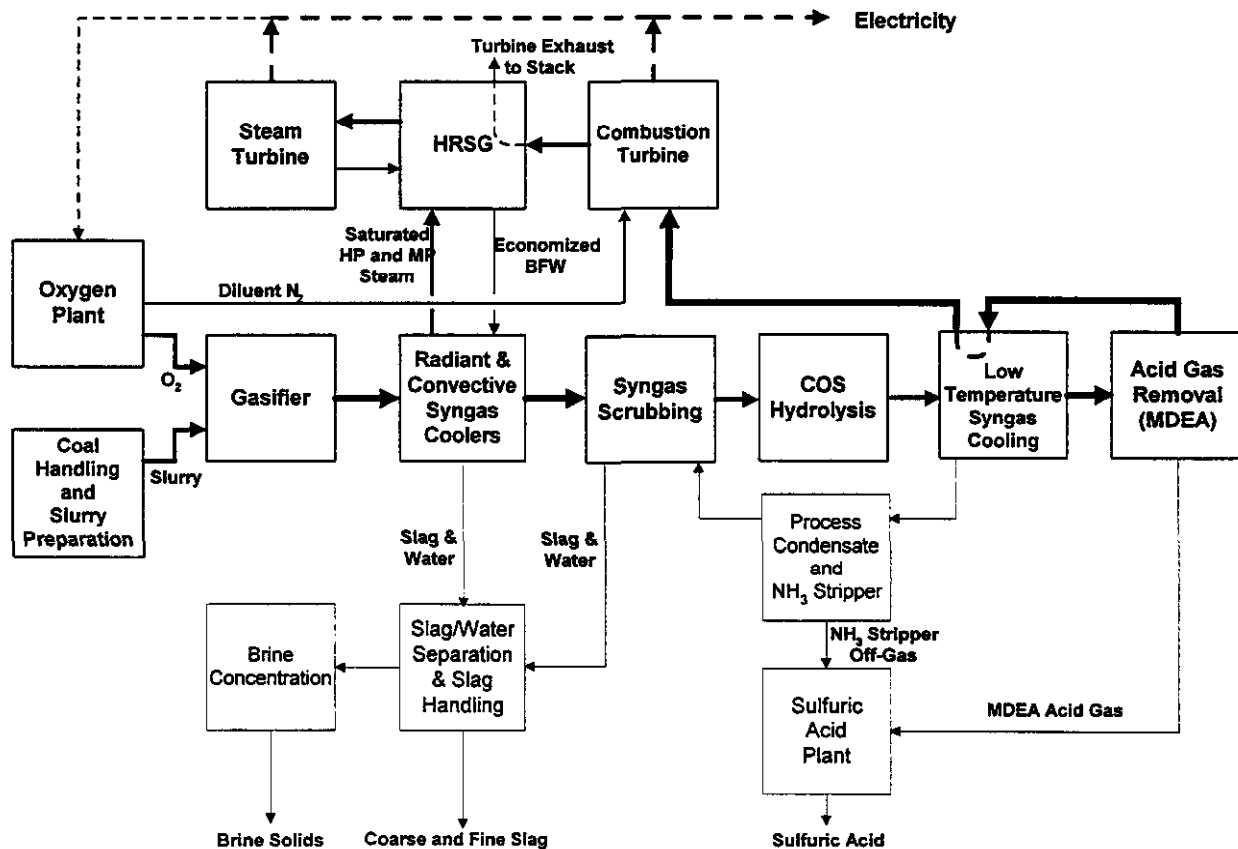
Polk Power Station is an integral part of TEC's generation expansion plan. TEC's original objective was to build a coal-based generating unit providing reliable, low-cost electric power. Integrated Gasification-Combined Cycle (IGCC) technology will meet those requirements.

Demonstration of the oxygen-blown entrained-flow IGCC technology is expected to show that such a plant can achieve significant reductions of SO_2 and NO_x emissions when compared to existing and future conventional coal-fired power plants. In addition, this project is expected to demonstrate the technical feasibility of commercial scale IGCC technology. Only commercially available equipment has been used for this project. The approach supported by DOE is the highly integrated arrangement of these commercially available pieces of hardware and systems in a new arrangement which is intended to optimize cycle performance, costs, and marketability at a commercially acceptable size of nominally 250 MW (net).

Technical Description

A general flow diagram of the entire process is shown in Figure 1.

FIGURE 1
Polk Unit #1 IGCC Block Flow Diagram



This unit utilizes commercially available oxygen-blown entrained-flow coal gasification technology licensed by Texaco Development Corporation (Texaco). In this arrangement, coal is ground with water to the desired concentration (60-70 percent solids) in rod mills. The unit is designed to utilize about 2200 tons per day of coal (dry basis). An Air Separation Unit (ASU) separates ambient air into 95% pure oxygen for use in the gasification system and sulfuric acid plant, and nitrogen which is sent to the advanced combustion turbine (CT). The ASU is sized to produce about 2100 tons per day of oxygen and 6300 tons per day of nitrogen. The ASU was provided by Air Products.

This coal/water slurry and the oxygen are then mixed in the gasifier feed injector. They react in the gasifier to produce syngas with a heat content of about 250 BTU/SCF (LHV). The gasifier is designed to achieve greater than 95 percent carbon conversion in a single pass. The gasifier is a single vessel feeding into one radiant syngas cooler (RSC) which was designed to reduce the gas temperature to 1400°F while producing 1650 psig saturated steam.

After the RSC, the gas is split into two (2) parallel convective syngas coolers (CSC), where the temperature is further reduced to less than 800°F and additional high pressure steam is produced. Next, the particulates and hydrogen chloride are removed from the syngas by intimate contact with water in the syngas scrubbers. The scrubbers are followed by a COS hydrolysis unit. The COS hydrolysis unit (now under construction) will enable Polk Power Station to continue processing high sulfur feedstocks and still meet the more stringent emissions restrictions which go into effect in late 1999. Following COS hydrolysis, most of the remaining sensible heat of the syngas is recovered in low temperature gas cooling by preheating clean syngas and heating steam turbine condensate. A final small trim cooler reduces the syngas temperature to about 100°F for the acid gas removal system.

The acid gas removal system is a traditional MDEA scrubber type which removes most of the sulfur from the syngas. This sulfur is recovered as sulfuric acid. The sulfuric acid plant was provided by Monsanto. Sulfuric acid has a ready market in the phosphate industry in the central Florida area.

Most of the residual solids from gasification fall into a water pool at the bottom of the RSC and then into the slag lockhopper which discharges them from the system. These residual solids generally consist of slag (the inert mineral matter from the feed coal) and some unreacted carbon. These non-leachable products are saleable for blasting grit, roofing tiles, and construction building products. TEC has been marketing slag from its existing units for such uses for over 25 years.

All of the water from the gasification process is cleaned and recycled, thereby creating no requirement for discharging process water from the gasification system. To prevent the build-up of chlorides in the process water system, a brine concentration unit removes them in the form of marketable salts.

The key components of the combined cycle are the advanced combustion turbine (CT), heat recovery steam generator (HRSG), steam turbine (ST), and electric generators. The combined cycle power block is provided by General Electric.

The CT is an advanced GE 7F machine adapted for syngas and distillate fuel firing. The initial startup of the power plant is carried out on low sulfur No. 2 fuel oil. Transfer to syngas occurs upon establishment of fuel production from the gasification plant. The exhaust gas from the CT passes through the HRSG for heat recovery, and leaves the system via the HRSG stack.

Emissions from the HRSG stack are primarily SO₂ and NO_x with lesser quantities of CO, VOC, and particulate matter (PM). SO₂ emissions are from sulfur species in the syngas which are not removed in the acid gas removal system. The CT uses nitrogen addition to control NO_x emissions during syngas firing. Nitrogen acts as a diluent to lower peak flame temperatures and reduce NO_x formation without the water consumption and treatment/disposal requirements associated with water or steam injection NO_x control methods. Maximum nitrogen diluent is injected to minimize NO_x exhaust concentrations consistent with safe and stable operation of the CT. Water injection is employed to control NO_x emissions when backup distillate fuel oil is used.

The HRSG is installed in the CT exhaust in a traditional combined cycle arrangement to provide superheated steam to the 130 MW ST. No auxiliary firing is done in the HRSG system. The HRSG high and medium pressure steam production is augmented by steam produced from the coal gasification plant's syngas coolers (HP and MP steam) and sulfuric acid plant (MP steam). All steam superheating and reheating is performed in the HRSG before the steam is delivered to the ST.

The ST is a double-flow reheat turbine with low pressure crossover extraction. The ST and associated generator are designed specifically for highly efficient combined cycle operation with nominal turbine inlet throttle steam conditions of approximately 1450 psig and 1000°F with 1000°F reheat inlet temperature.

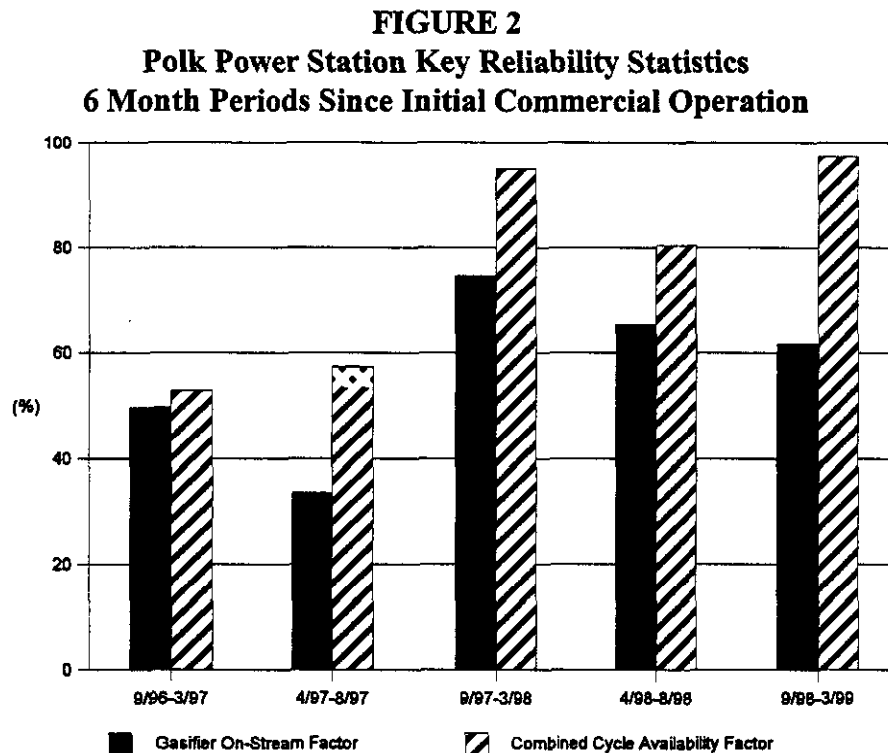
The heart of the overall project is the integration of the various pieces of hardware and systems to increase overall cycle effectiveness and efficiency. In our arrangement, benefits are derived from using the experience of other IGCC projects, such as the Cool Water Coal Gasification Program, to optimize the flows from different subsystems. For example, low pressure steam from the HRSG and extraction steam from the ST supply heat to the gasification facilities for process use. The HRSG also receives steam energy from the syngas coolers and sulfuric acid plant to supplement the steam cycle power output. This steam is generated using boiler feedwater which had been economized in the HRSG. Additional low energy integration occurs between the HRSG and the gasification plant. Condensate from the ST condenser is returned to the HRSG/integral deaerator by way of the gasification area, where condensate preheating occurs by recovering low level heat. Probably the most novel integration concept in this project is our use of the ASU. This system provides oxygen to the gasifier in the traditional arrangement, while simultaneously using what is normally excess or wasted nitrogen to increase power output and improve cycle efficiency and also lower NO_x formation.

Part of our cooperative agreement with DOE is a five-year demonstration phase. During the first two years of this period, several different types of coal or coal blends have been tested in the operating IGCC power plant. The results of these tests compare this unit's efficiency, operability and costs on each of these test coals against the design basis coal, a Pittsburgh #8. These results provide a menu of operating parameters and costs which can be used by utilities in the future as they make their selection on methods for satisfying their generation needs, in compliance with environmental regulations.

II. RELIABILITY GROWTH AND LOST PRODUCTION CAUSES

Overall

In its 2½ years of commercial operation, the Polk Power Station gasifier has operated over 13,000 hours. 3½ million MWH of electricity have been generated from the syngas fuel it produced. It was on-line 65% of the time for the last 1½ years. Even when the gasifier was unavailable, the combined cycle was available for operation on distillate fuel most of the time. The combined cycle's availability was 92% for the last 1½ years, and 99% for the six-month period ending April 1, 1999. These statistics are illustrated in Figure 2.



Gasifier on-stream factor and combined cycle availability were only about 50% during the first six months of commercial operation. This is to be expected with any new facility.

During the second six-month period of commercial operation, availability of the gasifier and combined cycle were both very low due primarily to two issues, both of which were discussed at length in last year's Clean Coal Technology Conference Paper. The worst of these was damage to the combustion turbine on two occasions from particulates in the syngas. On the first occasion, the particulates were coal ash from tube leaks in a raw gas/clean gas exchanger. On the second occasion, the particulates were primarily pipe scale from the syngas line. The problematic exchangers were removed and a filter has been installed immediately upstream of the turbine to catch the pipe scale. A recent hot gas path inspection of the turbine confirmed that these problems are behind us. The second significant issue during this 6 month period was seal leakage in the radiant syngas cooler. One seal was improperly manufactured, and it was repaired. Operating procedures were developed to deal with other smaller seal leaks. We believe that this issue also has been resolved.

During the third six-month period of commercial operation (October, 1997, through March, 1998) the station experienced excellent gasifier on-stream factor (75%) and combined cycle availability (95%). This was consistent with our expectations for this point in the plant's life-cycle.

The combined cycle continues to perform well, although its availability suffered slightly during the second and third quarters of 1998 (the fourth six-month period of commercial operation) due to a planned outage and steam turbine condenser tube leaks caused by human error. However, the gasifier's on-stream factor has only averaged about 63%, which does not meet expectations. In addition to two planned outages, the gasification system has experienced several forced outages. Three specific problems have had the greatest impact. They will be discussed in more detail in the next section of this paper. They are:

- a) Fuel changes causing slag removal and slurry feed problems,
- b) Raw syngas line leaks, and
- c) Convective Syngas Cooler pluggage

Recent Lost Production Causes April 1, 1998, Through June 15, 1999 *An Update Since Last Year's Clean Coal Technology Conference*

Gasification system lost production causes since our last Clean Coal Technology Conference Paper are summarized in Table 1.

A comparison of this year's list to last year's shows that although we have several new challenges, we were very successful in eliminating the worst causes of lost production from last year. Specifically, the top two causes of lost production from last year's list (particulate contamination of syngas to the turbine and RSC Dome Seal Leaks) do not appear on this year's list at all. These problems are behind us. Many other problems from last year's list have been eliminated or their impact dramatically reduced.

The recent lost production causes (our new challenges) are discussed in more detail in the remainder of this section of the paper.

<p>TABLE 1 Lost Production Summary - Gasifier Runs 51-99 April 1, 1998 - June 17, 1999</p>				
EVENT / CAUSE	FORCED GASIFIER OUTAGES/ STARTUP DELAYS	RESULTING GASIFIER OUTAGE DAYS	REMEDIAL ACTION	
PLANNED OUTAGES	3	58		
RAW SYNGAS PIPING EROSION Scrubber Overhead Scrubber Inlet	6 2	29 4	Harder Materials In Specific Locations Stop Running Scrubbers In Carryover Mode Some Minor Configuration Changes	
FUEL SWITCHING PROBLEMS Slag Removal Slurry Feed	7 2	22 5	Longer Runs on Consistent Fuel Slag Removal System Mods/Improvements Operating Experience	
CONVECTIVE SYNGAS COOLER PLUGGING	6	24	Modify Configuration Operating Procedures	
GASIFIER REFRACTORY REPLACEMENT	1	15	Longer Runs on Consistent Fuel	
RSC WATERWALL LEAK (Transmitter Failure)	1	7	Transmitters Replaced & Installed Properly	
CONDENSER TUBE LEAKS (Controls/Fuse)	1	5	Improved Procedures/Controls	
SLURRY FEED PUMP	1	5	Improved Materials	
BLACK WATER PIPING LEAKS	2	4	Improved Materials & Configuration Changes Routine Inspection	
MISCELLANEOUS FORCED OUTAGES, OUTAGE EXTENSIONS, AND MAINTENANCE OUTAGES Oxygen Plant Gasification Power Block/Transmission System/Utilities	4 17 8	3 6 4	Various Mechanical Improvements, Controls Modifications, and Procedural Changes	

Planned Outages: 3 Occurrences, 54.4 Days Gasification Lost Production

Polk Power Station expects to have one major planned outage per year of 20 to 30 days' duration. Both the 1998 and the 1999 planned outages fell within this most recent reporting period.

The 1998 annual planned outage occurred in April/May. Both the power block and gasification system were unavailable for 31 days.

The 1999 gasification and power block outages were to have been combined as they were in 1998. However, when the gasifier came off-line two days early due to an oxygen valve positioner failure, it was decided to keep it off-line and get a head start on the gasification outage. The power block was to remain available for two more days to meet system demand. Unfortunately, system demand remained high throughout the entire gasification outage, so we were not able to do the necessary power block maintenance during that period. The power block outage was taken two weeks later. Consequently, the 1999 planned outages totaled 23.6 days, about twice as long as expected.

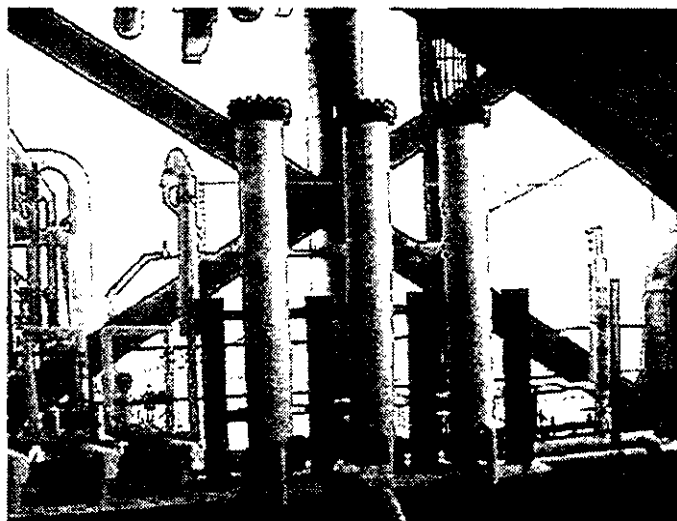
Syngas Syngas Scrubber Overhead Piping: 6 Occurrences; 29 Days Gasification Lost Production

Polk's Texaco gasification system produces more than expected of one specific sulfur compound, carbonyl sulfide or COS. Our acid gas removal system, MDEA, does not remove COS from the syngas, so any COS produced is converted to SO₂ emissions in the HRSG stack. This was not a problem with the relatively expensive design coal, Pittsburgh #8, since it only contained about 2.5% sulfur. However, in an effort to reduce the cost of electricity for our ratepayers and to meet DOE requirements, we began testing various less expensive feedstocks such as Illinois #6 and Kentucky #11 coals with sulfur content up to 3.5%. These higher sulfur coals produced proportionally more COS, so our SO₂ emissions from the HRSG stack would have exceeded our permit limits except for one factor: we discovered a method during early operation to reduce the COS content of the syngas by about 30%. Specifically, by flooding the syngas scrubber overhead lines with particulate laden water, about 30% of the COS is converted (hydrolyzed) to H₂S as the syngas passes through them. This probably occurs by the reaction: $\text{COS} + \text{H}_2\text{O} \rightarrow \text{CO}_2 + \text{H}_2\text{S}$. Flooding the scrubber overhead lines to reduce COS enabled us to operate on the higher sulfur less expensive Illinois #6 and Kentucky #11 seam coals from mid-December, 1997, until mid-November, 1998, without exceeding our emissions permits. See US Patent Application 60/112,335.

Flooding the scrubber overhead piping had one significant drawback: the scrubber overhead piping system was not designed for this turbulent three-phase operation. After only four months of operation in this manner, we experienced the first very small localized syngas leak. The first leaks were pin-holes which could be easily repaired. With each repair, we reinforced and/or hard surfaced the damaged area and intensified testing. The damage was very localized, so conventional ultrasonic testing was not completely effective in identifying damaged areas. In response, we developed an improved testing technique. We also began planning replacement of the piping system and accelerated our plans to install a conventional COS hydrolysis unit. Finally, in November, 1998, a larger leak occurred which prompted us to take a 23 day forced outage to replace the entire piping system with the upgraded materials. We also decided to no longer operate with the piping system flooded (even though the new piping system could probably accommodate it). Instead, we elected to process lower sulfur coals until COS hydrolysis became available in September, 1999.

The process design for the COS hydrolysis was done at Polk by TEC personnel. Catalyst selection and reactor sizing was the result of the testing we performed on the actual raw syngas stream. Three parallel test reactors were used so competing catalysts could be compared side-by-side under the same actual plant conditions. The test reactors are shown in Figure 3.

Figure 3
COS Hydrolysis Test Reactors



Fuel Related Problems: 9 Occurrences; 27 Days Gasification Lost Production

The need to temporarily gasify lower sulfur coals in response to the problem with the syngas scrubber overhead piping system prompted us to test some blends of higher sulfur Kentucky #11 (our base fuel) with lower sulfur fuels to produce a reasonable cost feedstock with an average sulfur content of about 2½%. These blends introduced unexpected problems during this reporting period. On 7 occasions, large slag agglomerates plugged the slag removal system at the bottom of the radiant syngas cooler, and on 2 occasions, slurry solids from these blends settled in the gasifier feed pump suction line, starving the pump. Together, these problems resulted in 27 days of lost gasifier production. During the previous reporting period, fuel changes led to three gasifier forced outages and 12 days of lost gasifier production.

There is no single simple solution to problems resulting from fuel changes. We have made modifications to the mechanical and process configuration of the slag removal system to help cope with large slag agglomerates. We identified a problem with one of the outside coal testing laboratories that provided critical analytical data on fuel shipments. Texaco is helping us apply some complimentary methods of characterizing slag viscosity. The ultimate solution to these difficulties probably will only come when we find a suitable feedstock and gain longer term operating experience on it.

Convective Syngas Cooler Plugging: 6 Occurrences; 24 Days Gasification Lost Production

Polk Power Station has horizontal fire-tube convective syngas coolers at the exit of the radiant syngas cooler. During the current reporting period, we experienced 6 outages and 24 total days of lost gasifier production due to pluggage of these exchanger tubes. This was not identified as a cause of lost production in previous reporting periods because we had opportunities to clean out the incipient deposits more frequently for other reasons. Purging, cooling, cleaning, reheating, and restarting take 3 to 4 days.

This pluggage occurs via two mechanisms. First, large ash agglomerates spall from the inlet duct which instantaneously plug several tubes. This has been predominate at startup. Modified startup procedures appear to mitigate this source of plugging. The second is the gradual build-up of deposits during operation. Ash constituents of some of our fuels may accelerate this pluggage. Some recent configuration changes seem to help reduce this pluggage rate.

We monitor the progression of the pluggage with the differential pressure across the exchangers. Significant tube damage occurred on one occasion due to erosion from the pluggage, and some minor damage occurred on another occasion, so we know approximately how much pluggage we can tolerate before we must shut down for cleaning. We can now operate 25 to 40 days between cleanings. We hope that our recent changes in configuration and operating procedures will enable us to operate 45 to 60 days or more between cleanings. If not, other options are being evaluated.

Gasifier Refractory Replacement: 1 Occurrence, 18 Days Gasification Lost Production

During the 1999 Power Block outage, routine inspection revealed that a section of the gasifier refractory had failed during the previous shutdown, so gasifier startup was delayed an additional 2½ weeks to rebrick the gasifier. That liner (Polk's first high quality liner) had survived 451 operating days of service across 755 calendar days on 10 different coals and/or blends through 74 startups. We had hoped it would last until fall of 1999, even though we knew it was severely worn. Nevertheless, its performance was satisfactory considering the service it had seen. We have every reason to expect at least 50% longer refractory life if we can find an economically attractive consistent feedstock and extend the mean gasifier run length.

Other Causes of Forced Outages and Lost Gasifier Production

There were a number of problems which altogether cost 38 days of lost gasifier production in this reporting period. Five of these cost between 4 and 7 days each. The remaining 29 cost a total of 13 days. All were one or two of a kind incidents. Some of those, including those with the greatest impact were:

RSC Waterwall Leak Flow transmitters monitor cooling water flow to a few critical radiant syngas cooler waterwall sections. We experienced one waterwall leak due to a complicated sequence of events initiated by a failure of one of these transmitters.

Scrubber Inlet Piping Erosion The syngas scrubber inlet piping system experienced some deterioration due to simple erosion by the dry particulate laden syngas from the syngas coolers. As with the scrubber overhead piping, most erosion was localized at 90 degree bends and branches. All vulnerable areas of this piping system were upgraded in the 1999 gasification planned outage.

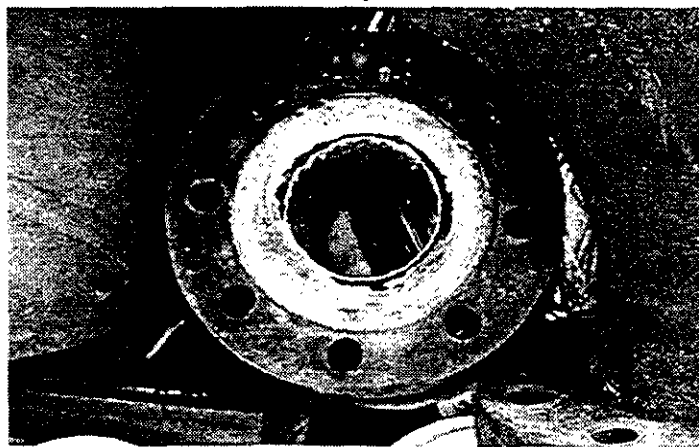
Steam Turbine Condenser Tube Leaks On one occasion, we lost cooling water flow to the steam turbine condenser during operation. This was a direct result of maintenance in progress on the control system, but a contributing factor was a weakness in the configuration of the control system for the circulating water pumps. Control system maintenance procedures and the control configuration were modified as a result of this incident.

Slurry Feed Pump Erosion/Corrosion The gasifier's slurry feed pump failed on one occasion during operation due to internal erosion/corrosion of the parts in contact with the coal/water slurry. Relatively high rates of metal loss in these parts had previously been noted. These parts were replaced with components fabricated with more corrosion-resistant materials. This has at least tripled the life of these parts. Regular maintenance and parts replacement will still be required on the pump every 1 to 3 months, but the alarmingly high rates of metal loss have been eliminated.

Black Water Piping Erosion The third most troublesome problem from last year's list was black water piping erosion which caused 8 forced outages resulting in 15 days of lost production in that reporting period. There was significant improvement since then. Although we did suffer two forced outages due to black water piping erosion since last year's conference, they only cost 4 days of lost production. The most recent of these occurred 10 months ago, in September, 1998, and the piping in this area has subsequently been hardened and modified. Although we can expect occasional forced outages in the future due to grey/black water piping erosion, this problem area seems to be under control.

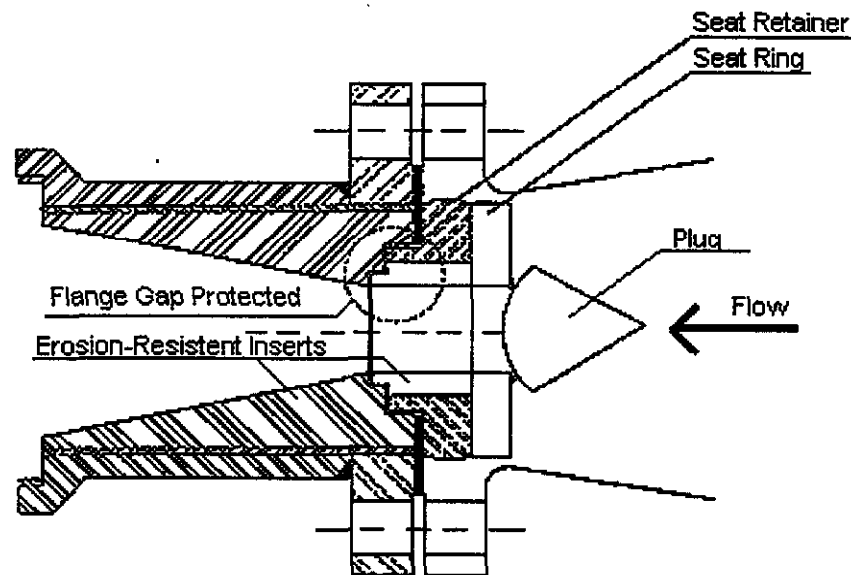
Each black water piping erosion failure is localized and has a relatively unique and sometimes interesting mechanism. For example, when the piping downstream of certain flow control valves exhibited high erosion rates, our first response was to coat the line and valve body with erosion resistant material. We found the erosion to be so persistent that it penetrated the joint between the flanges connecting the valve with the downstream piping and undercut the hard facing (Figure 4). This resulted in almost as rapid a failure as before.

FIGURE 4
Black Water Erosion Penetrates Flange Joints and Undercuts Hard Facing



The solution was to design and machine an assembly which consisted of the downstream piping spool and an insert to bridge the gap between it and the valve body (Figure 5).

FIGURE 5
Erosion Resistant Valve Assembly: Valve, Flange Gap Insert, and Downstream Spool
US Patent Application 09/243,331



Miscellaneous Minor Lost Production Causes 29 incidents occurred which, altogether, cost 13 days of lost gasifier production. 6 were in the air separation unit, 8 resulted from power block problems, and the remaining 17 were in the gasification area. Three were gasifier trips resulting from various upsets which led to a low level in the syngas cooler medium pressure steam drum. This drum's level indication and control is relatively unstable. The number of forced outages due to transmission system voltage swings and slag crusher seal failures has been cut in half (to one each) in the current reporting period compared to last year due to the successful remedial action we have taken. We still expect occasional forced outages due to these problems, but they seem to be under control. We have taken appropriate corrective action wherever practical in this "miscellaneous" category, so we expect continued improvement. However, such sources of lost production can never be entirely eliminated.

III. PLANS FOR 1999-2000

The following are some of the significant activities planned for Polk Power Station for the remainder of 1999 and into 2000.

1. Complete construction and commission the **COS hydrolysis** unit in September, 1999. We expect that by the beginning of 2000, the COS hydrolysis unit will enable us to settle in with a consistent, economical base feedstock which will reduce forced outages

due to fuel switching. It will also enable us to increase byproduct sales revenues by operating on higher sulfur fuels while staying well within SO₂ emission limits.

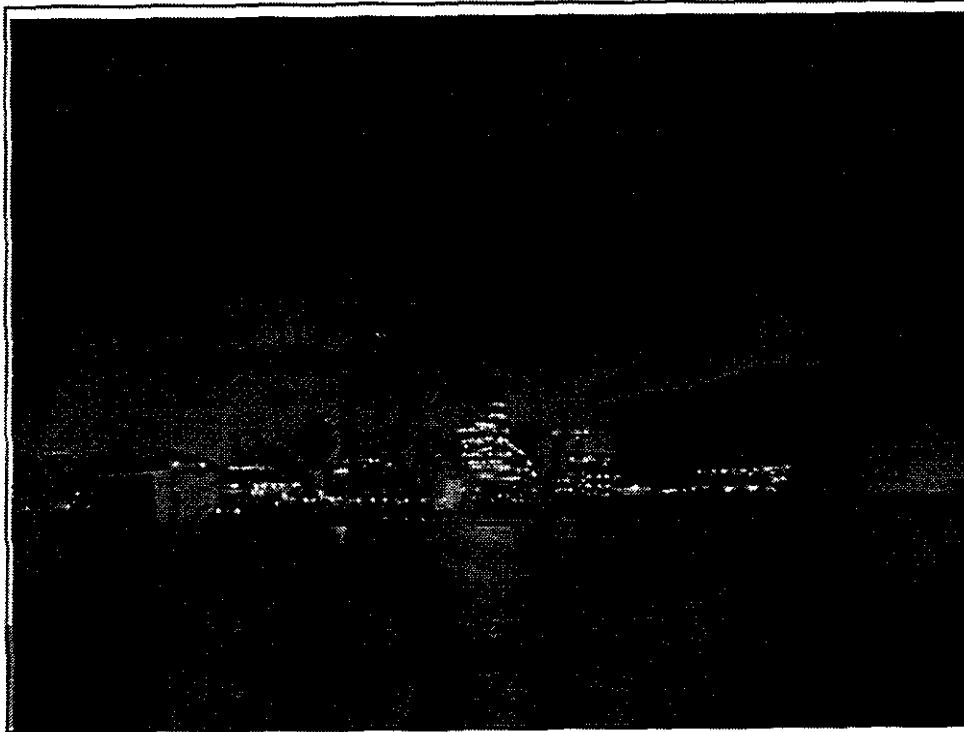
2. Continue efforts to reduce/eliminate **convective syngas cooler tube pluggage**.
3. Upgrade the **brine concentration system** to improve its reliability and lower overall plant heat rate. The majority of the brine concentration originally was accomplished via a vapor compression cycle. However, the vapor contained sufficient corrosive brine to make the compressor/blowers inoperable. Since the blowers have deteriorated, the system has been operated as a much less efficient direct evaporation system using low pressure steam. In 1999, a vapor scrubber will be added. We expect this to clean the vapor so a blower can function. Once this is demonstrated, a new blower will be added in 2000 to return brine concentration to its original vapor compression configuration.
4. Upgrade the **slag handling system** to reduce O&M costs, to produce a more valuable byproduct slag, and to enable selective recycling of some fractions of the current slag product to reduce heat rate. The design for the revised system was based on the alternate fuel test results to date. Detailed design for these design revisions has been completed and much of the equipment has been purchased. The system will be installed in 2000.
5. Identify a consistent, economical base coal for normal operation and continue selective **testing of alternate fuels** to lower Polk Power Station's overall busbar cost. We expect to begin initial testing of some petroleum coke blends in late 1999 as soon as COS hydrolysis has been successfully commissioned.

IV. CONCLUSIONS

Polk Power Station's leading causes of lost production in late 1997/early 1998 were particulate contamination of the clean syngas to the turbine and radiant syngas cooler seal leaks. These were successfully eliminated in the most recent 15 month period of operation, and many other less significant problems were either entirely eliminated or significantly reduced. However, performance in the last half of 1998 and the first half of 1999 was adversely impacted primarily by two new issues,

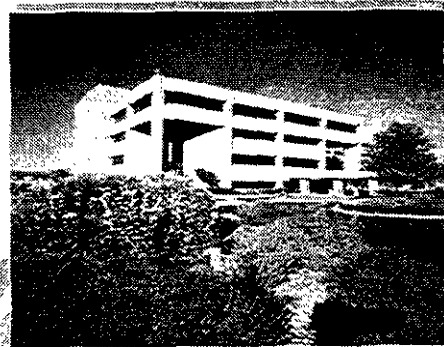
- a) syngas scrubber overhead line erosion and fuel switching resulting from dealing with higher than design carbonyl sulfide (COS) production, and;
- b) convective syngas cooler tube plugging.

The COS issue will be resolved with the commissioning of a COS hydrolysis unit in September, 1999. Steps are being taken to deal with the CSC pluggage, but that outcome is less certain. Nonetheless, we expect significant improvement over the next 12 month period. This should bring us closer to reaching our ultimate commercial goals in the areas of high reliability and efficiency with low emissions and busbar cost.



Sierra Pacific Power Company


- ◆ Investor-owned utility providing electric, gas and water services to 50,000 square miles of Northern and Central Nevada, and Northeastern California
- ◆ Our Corporate Headquarters are in Reno, NV
- ◆ Since '94 Sierra has invested \$250 million in 3 power projects




New Power Generation

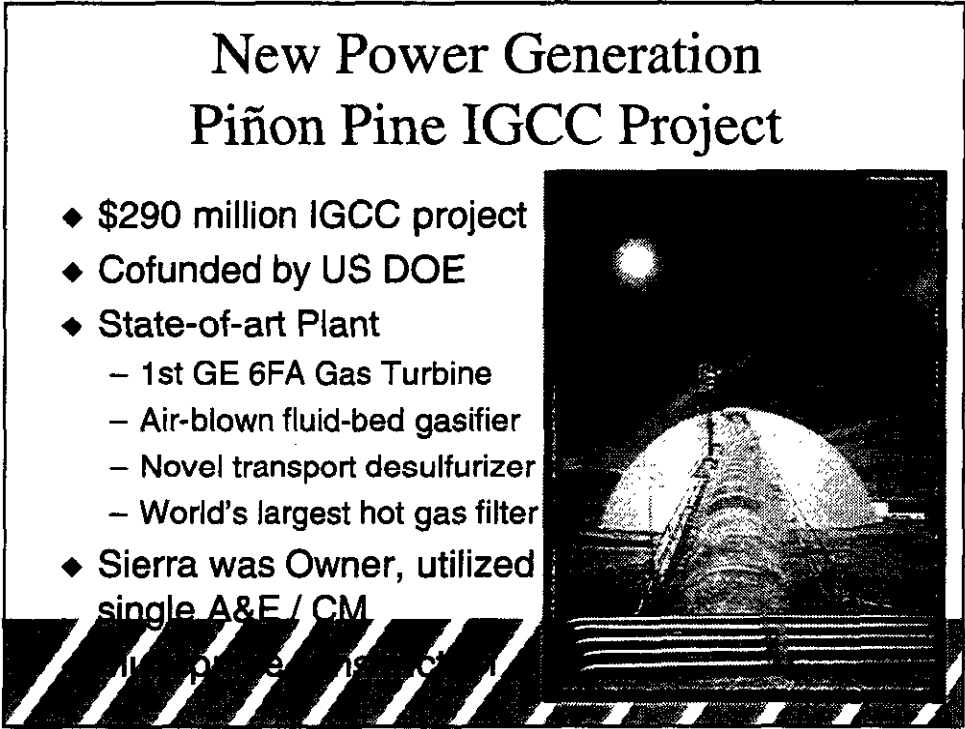
Piñon Pine IGCC Project

- ◆ \$290 million IGCC project
- ◆ Cofunded by US DOE
- ◆ State-of-art Plant
 - 1st GE 6FA Gas Turbine
 - Air-blown fluid-bed gasifier
 - Novel transport desulfurizer
 - World's largest hot gas filter
- ◆ Sierra was Owner, utilized single A&E / CM



Sierra was Owner, utilized single A&E / CM

- # New Power Generation
- ## Piñon Pine IGCC Project
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- 
- Sierra was Owner, utilized single A&E / CM



Block Diagram

Combined Cycle

Offsites

Denotes LLC

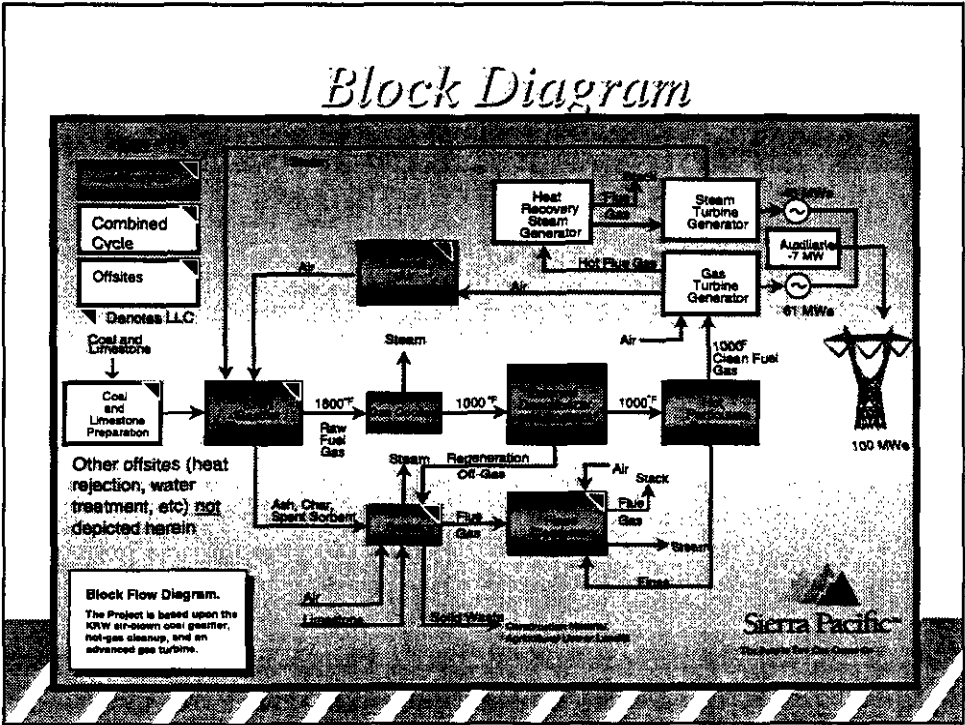
Coal and Limestone Preparation

Other offsites (heat rejection, water treatment, etc) not depicted herein

Block Flow Diagram.

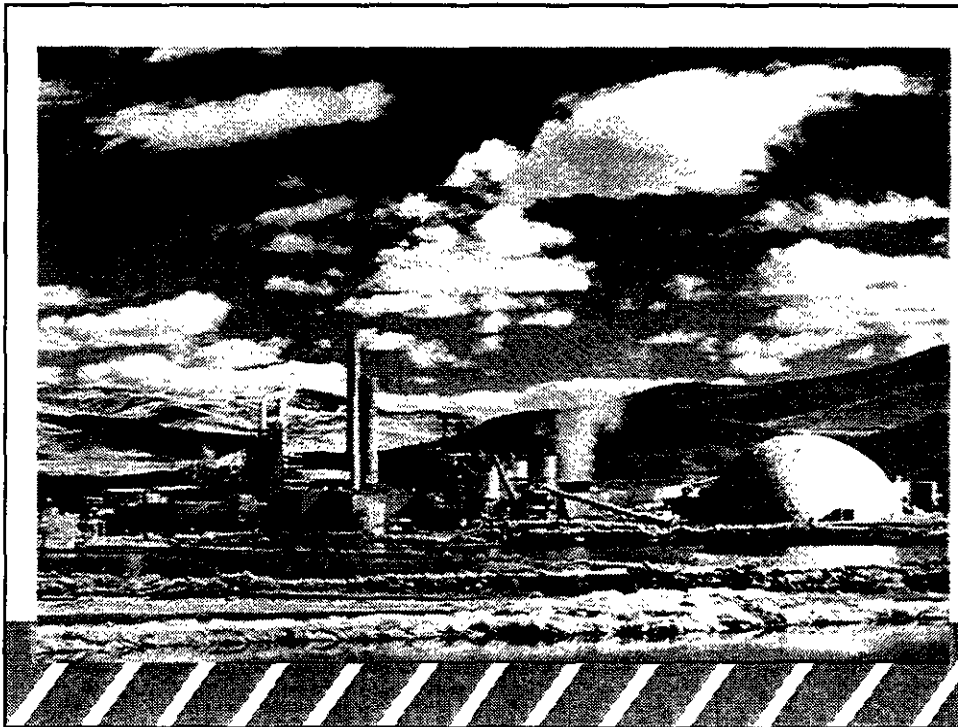
The Project is based upon the KRW air-blown coal gasifier, hot-gas cleanup, and an advanced gas turbine.

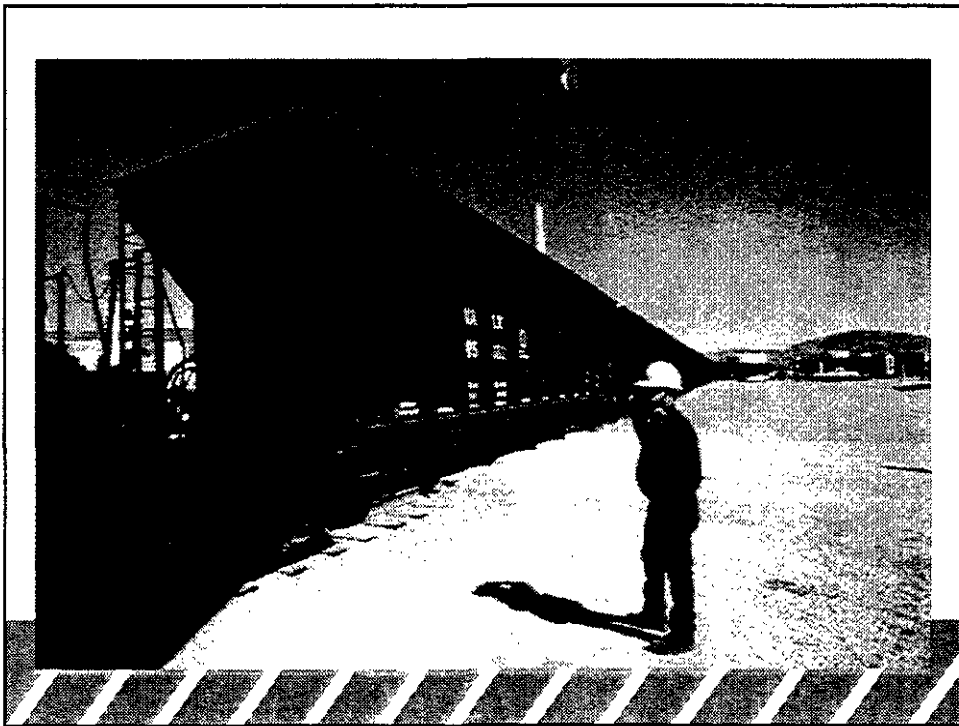
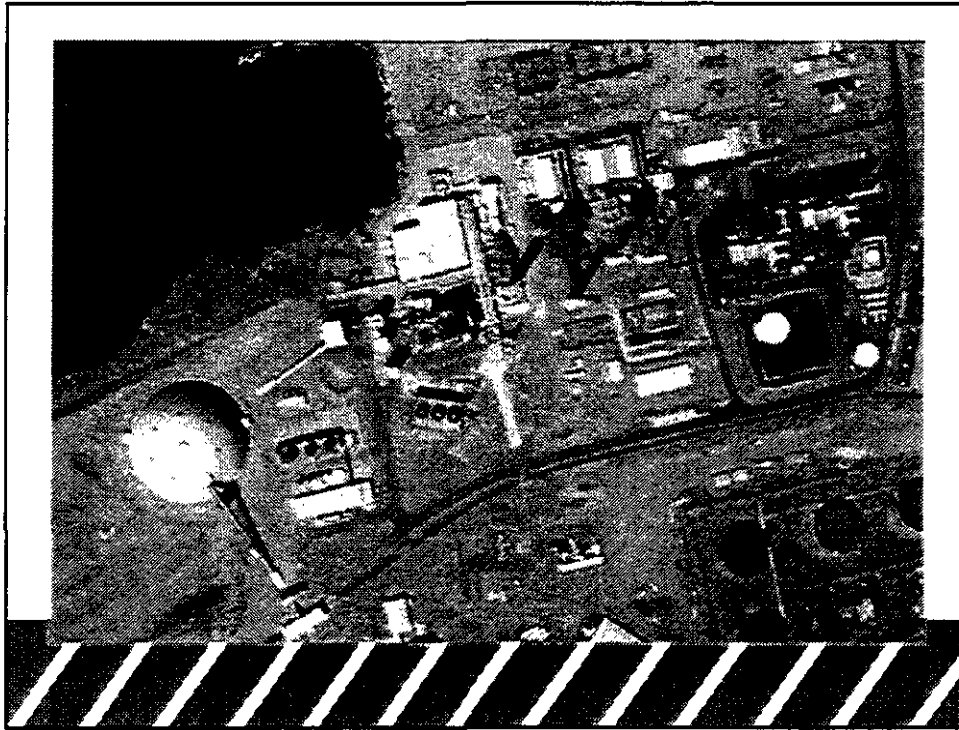
Sierra Pacific
The Nation's Best Gas Owner

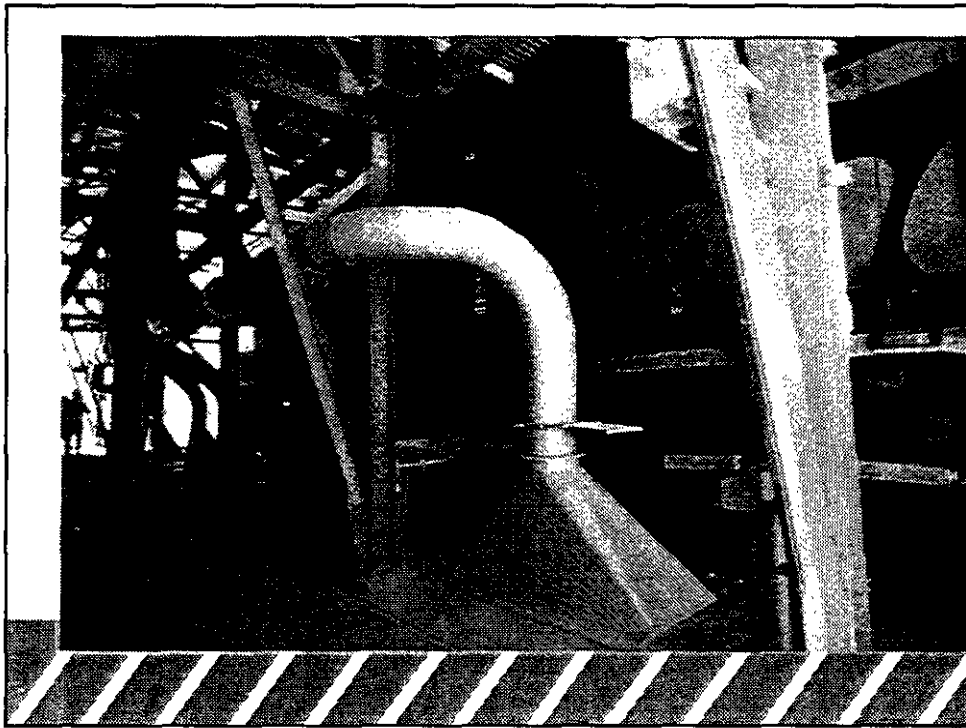


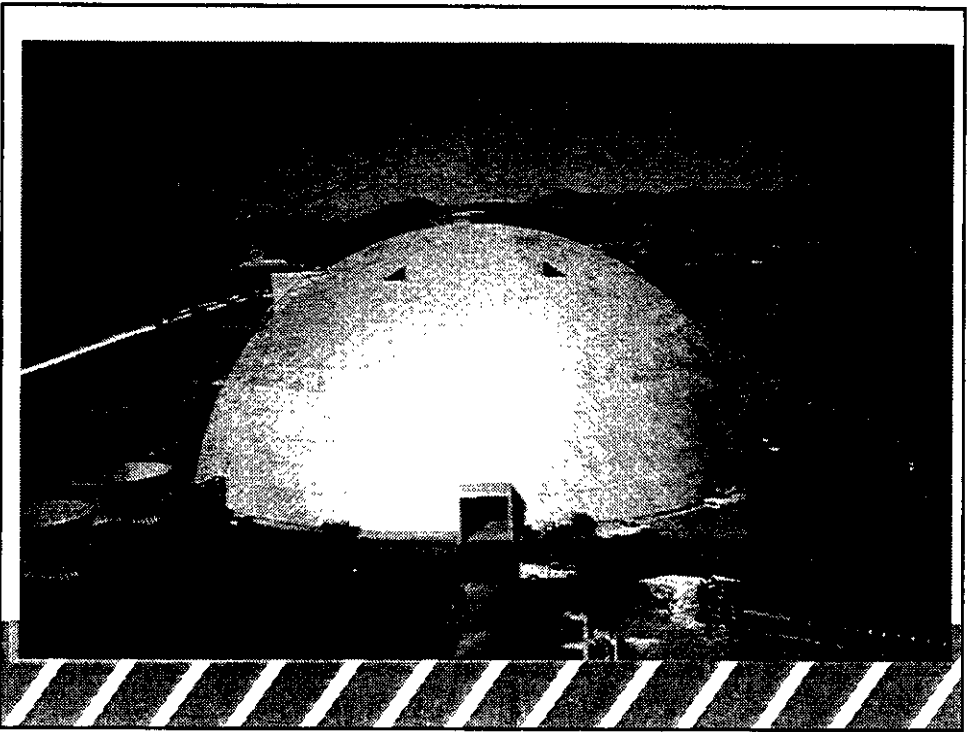
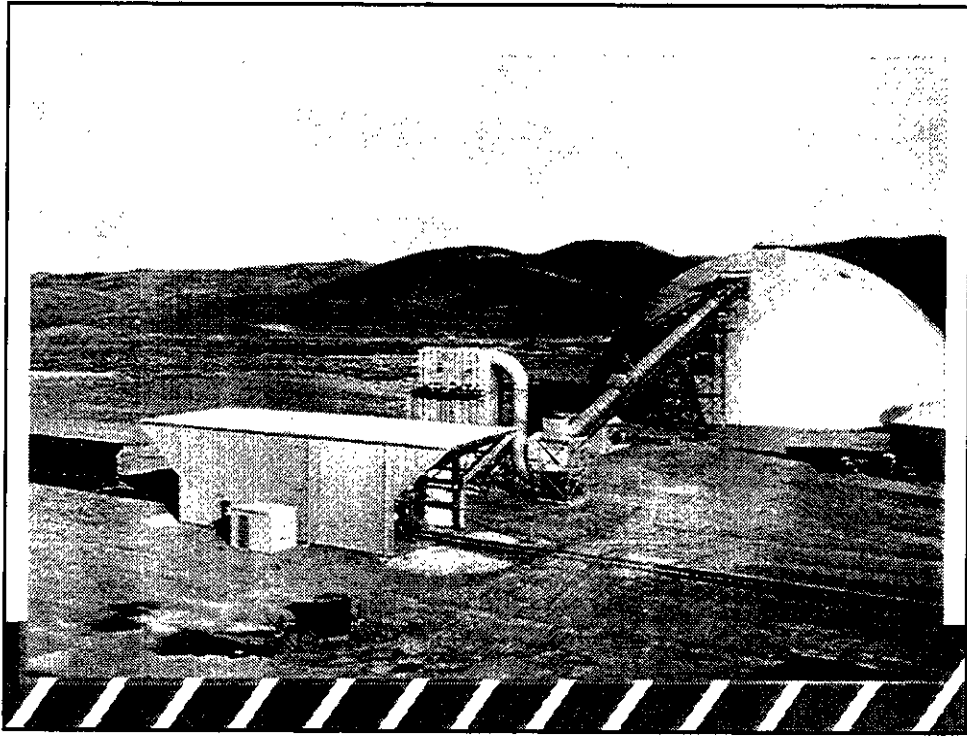


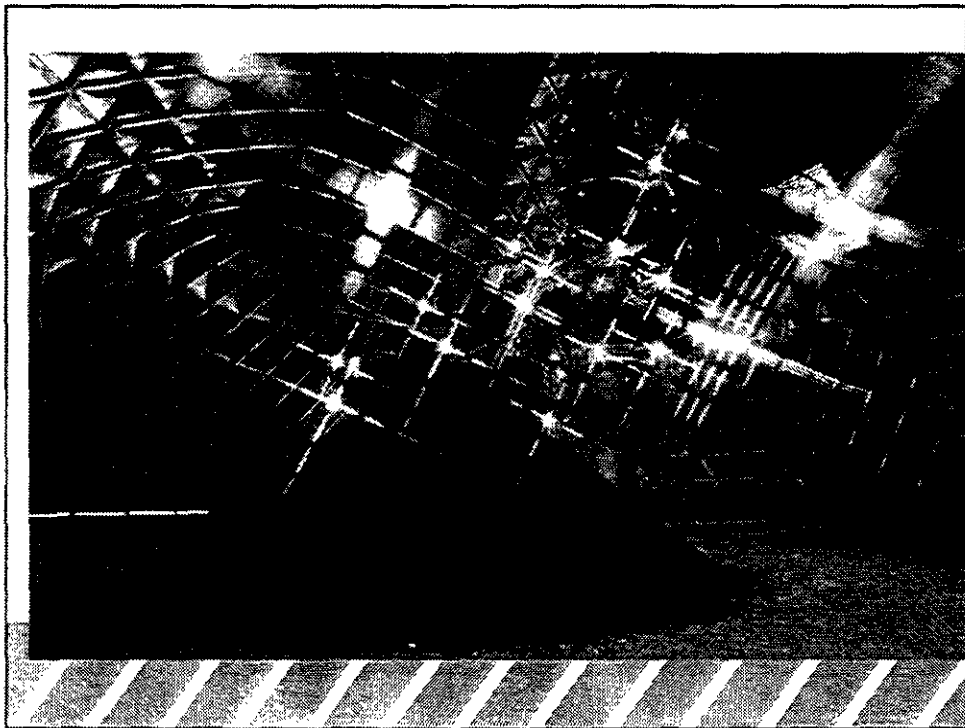
Photographs of Plant

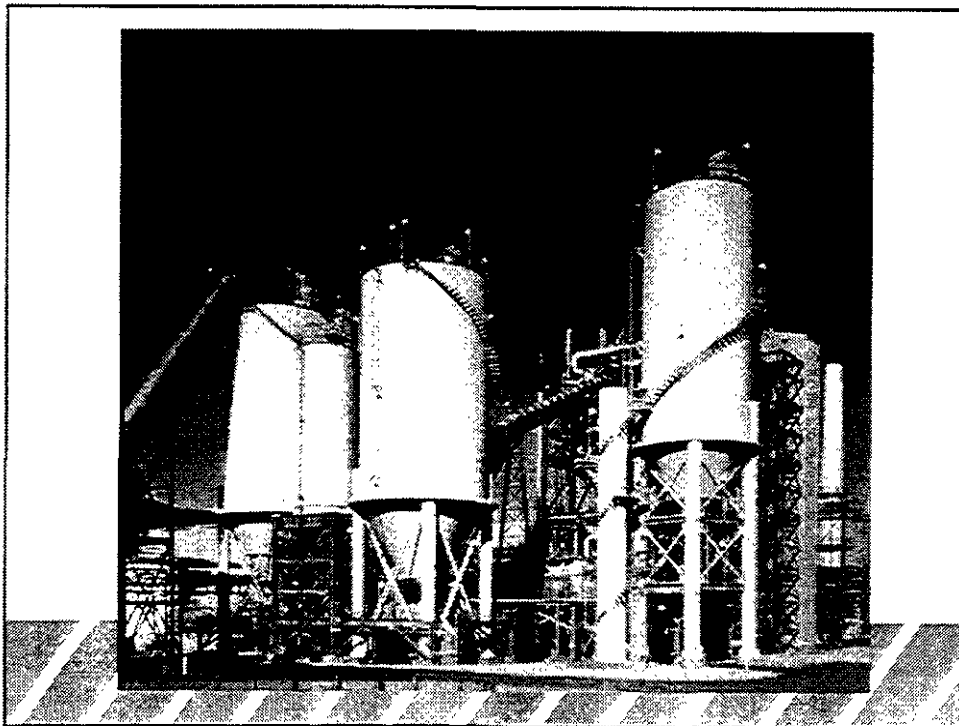
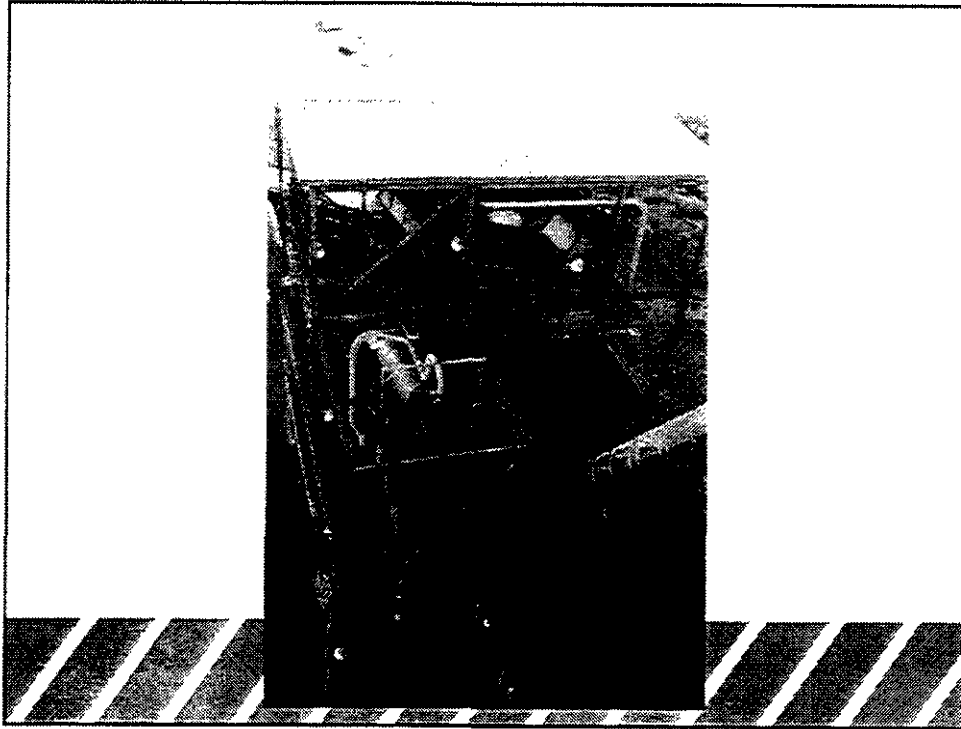


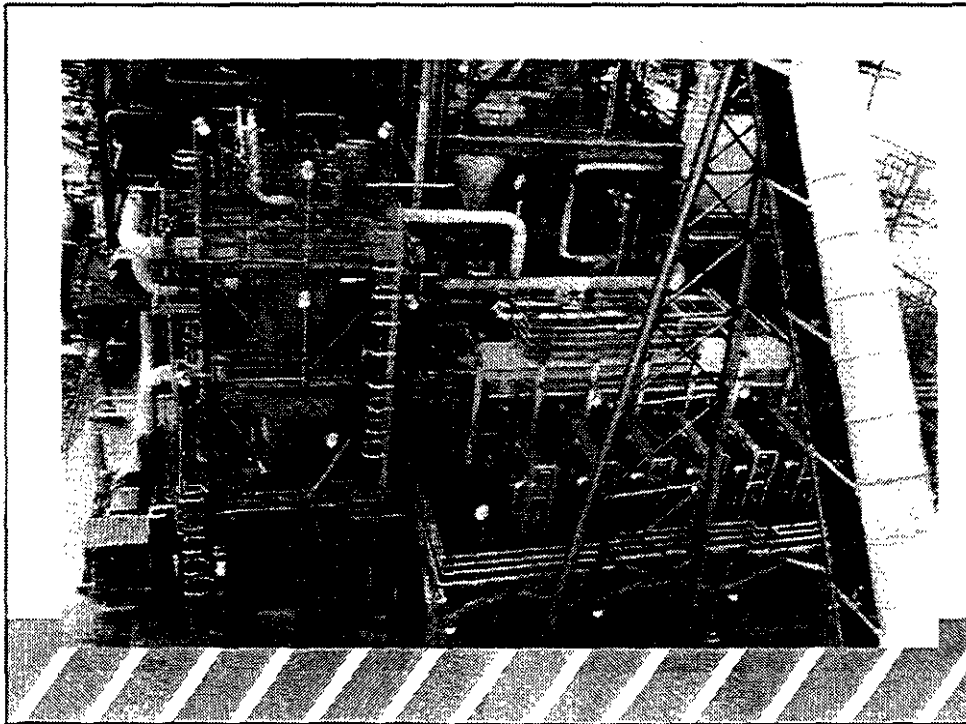
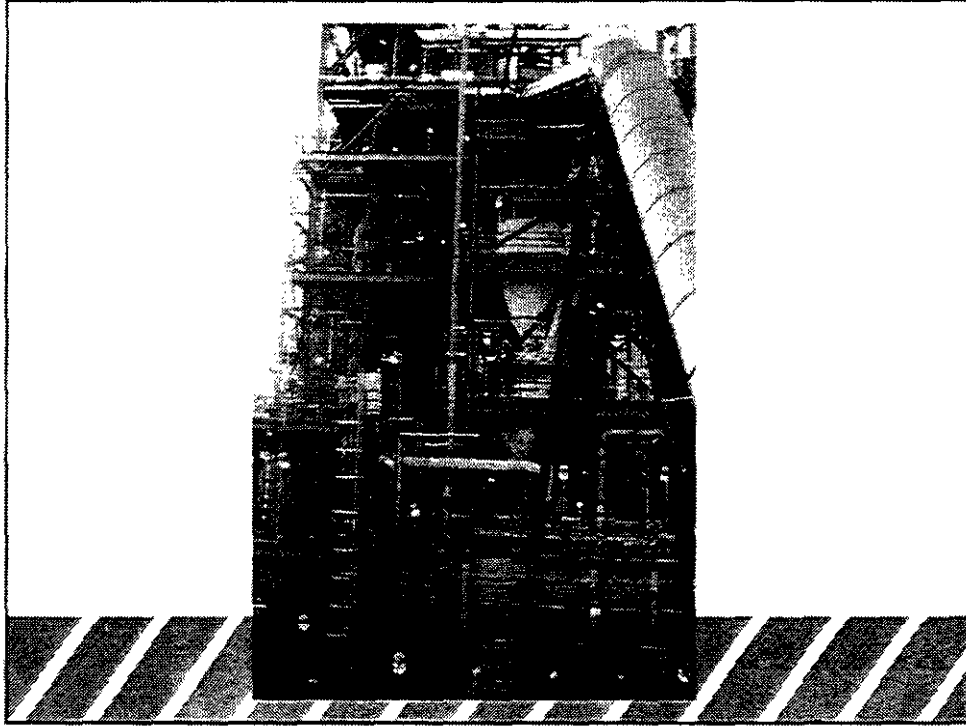


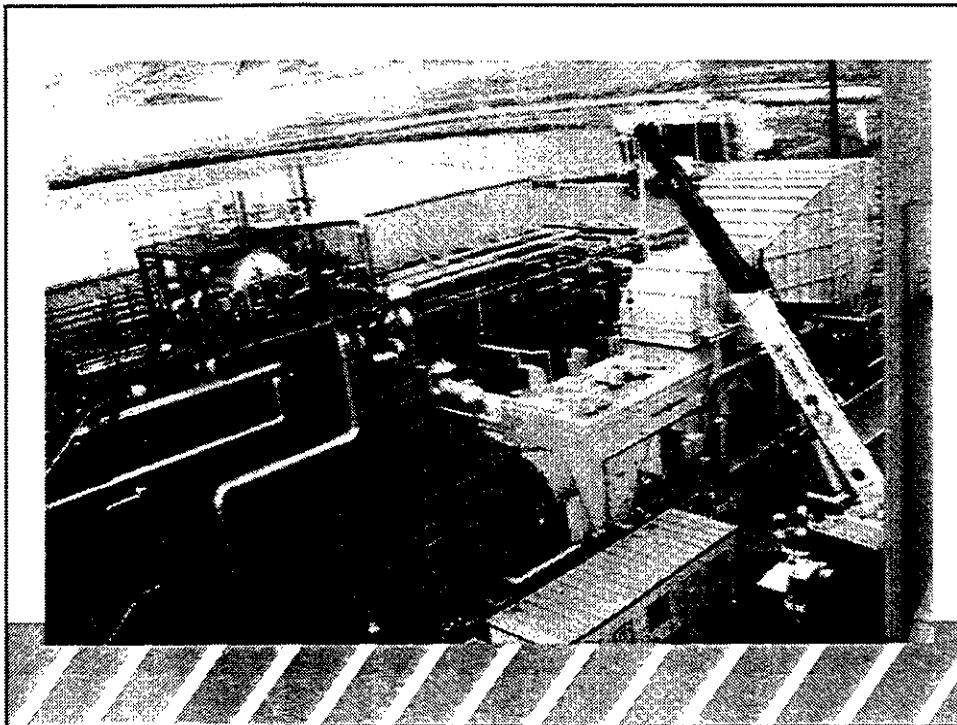


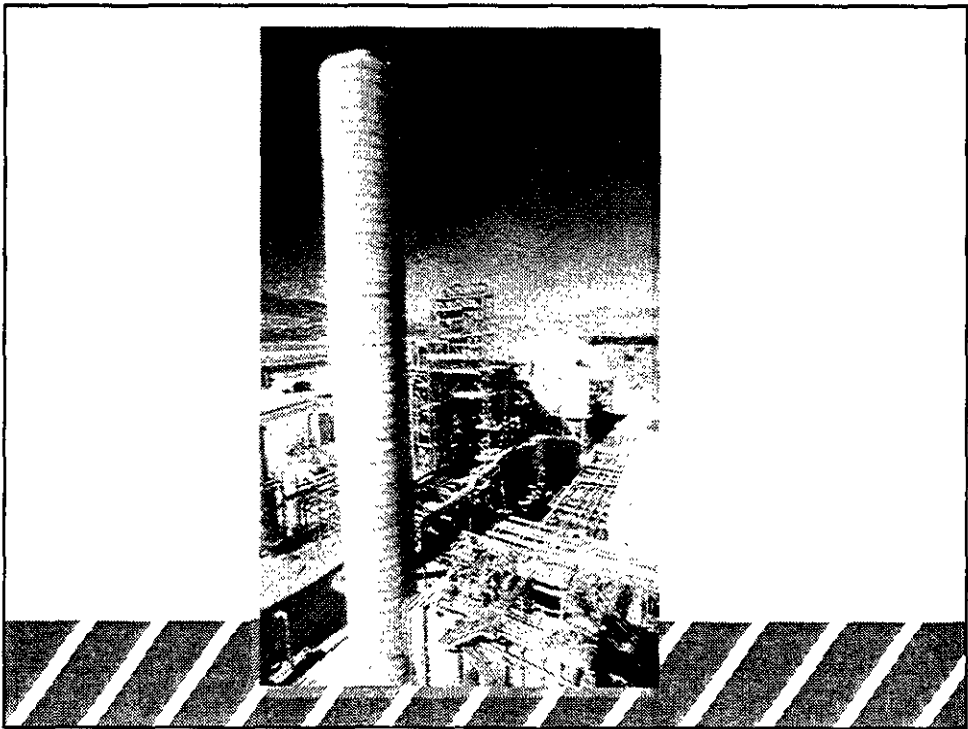
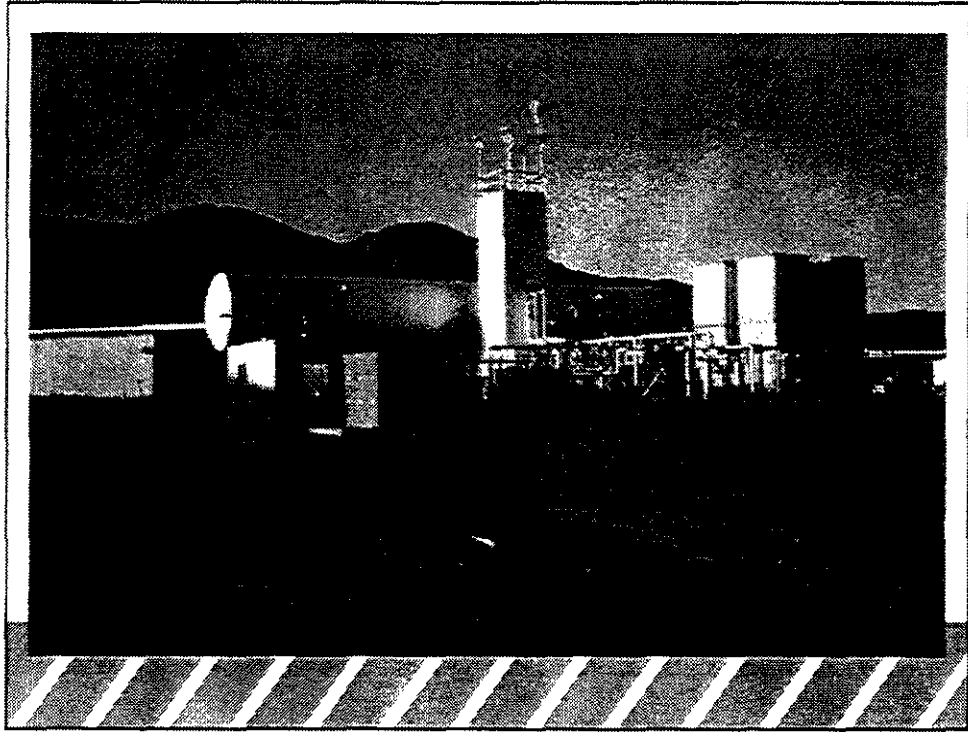


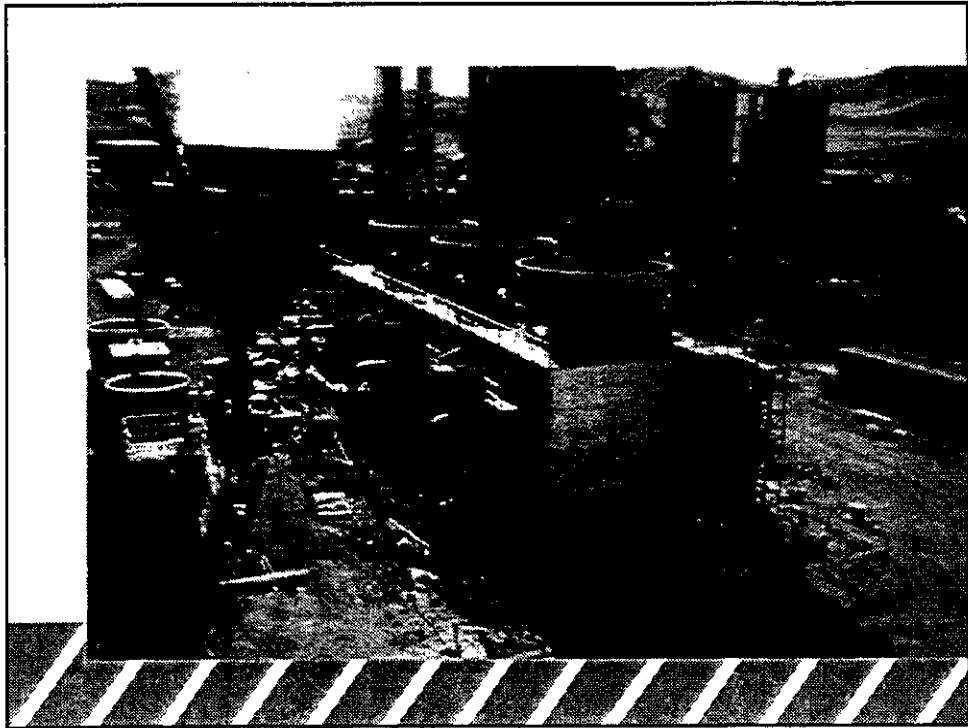
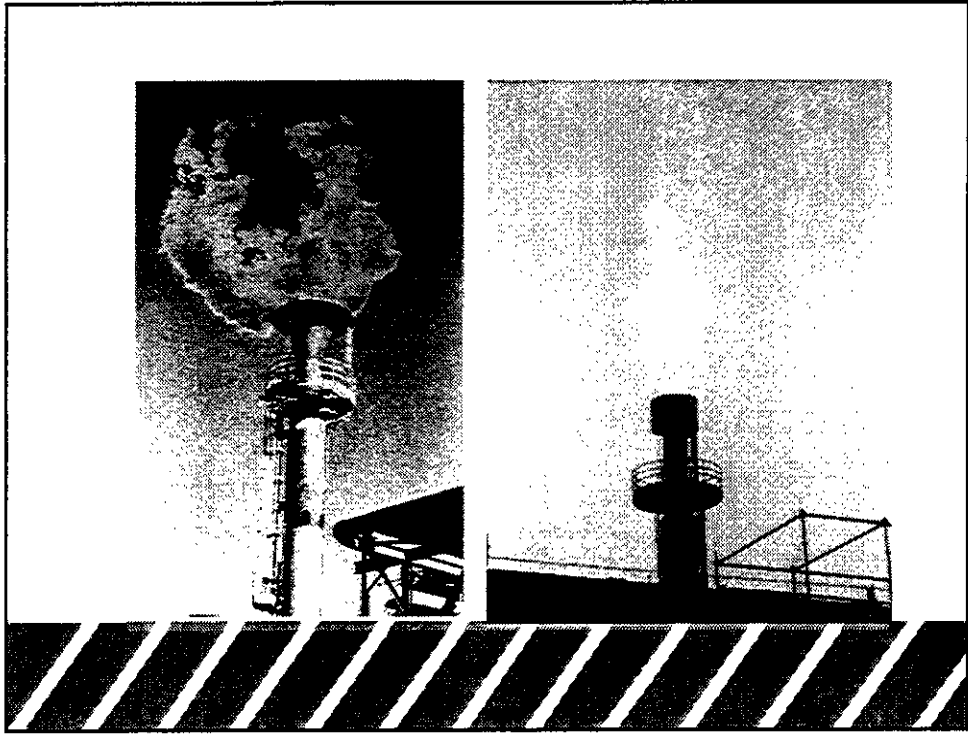


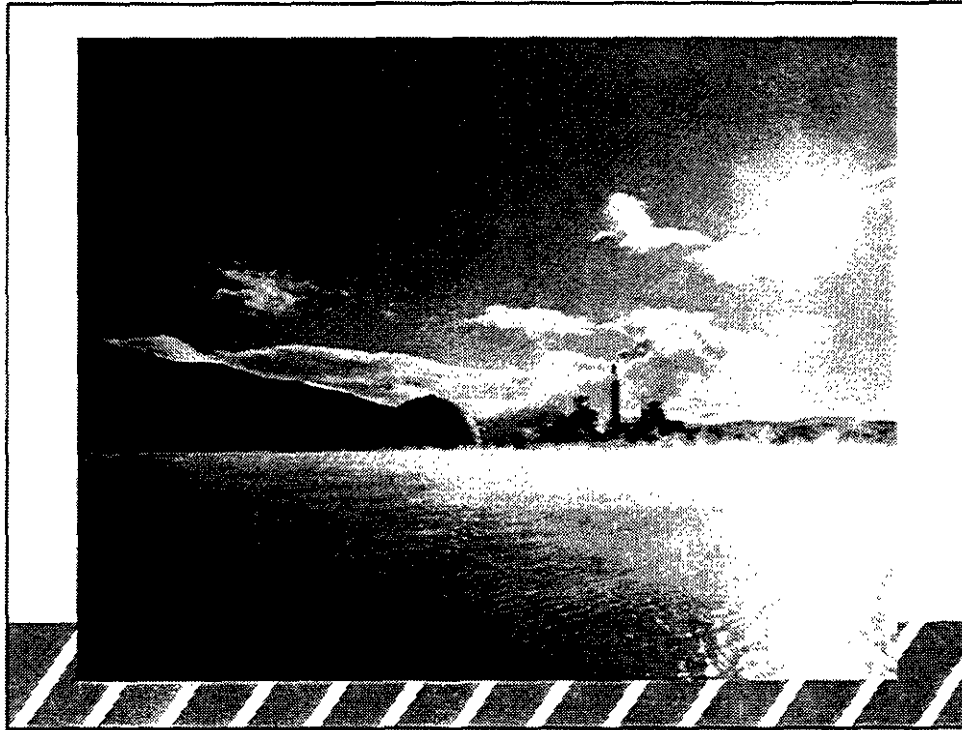












1998-99 Plant Performance

✓ Combined Cycle Performance

- 85% availability in 1998 (due to GT outages)
- 100% in 1999 to date

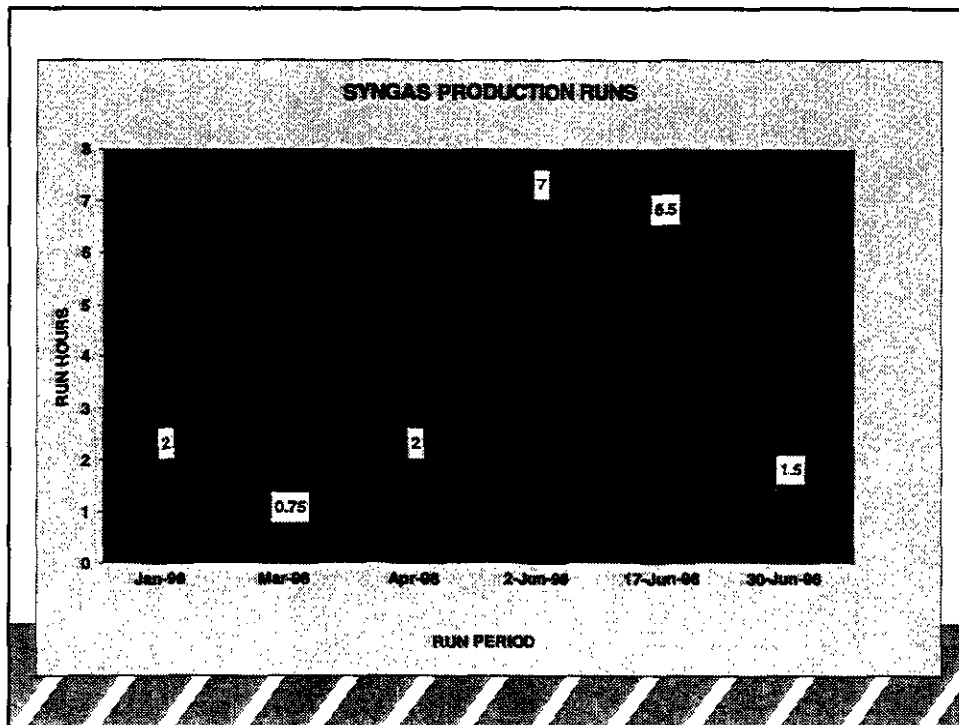
✗ Gasifier Performance

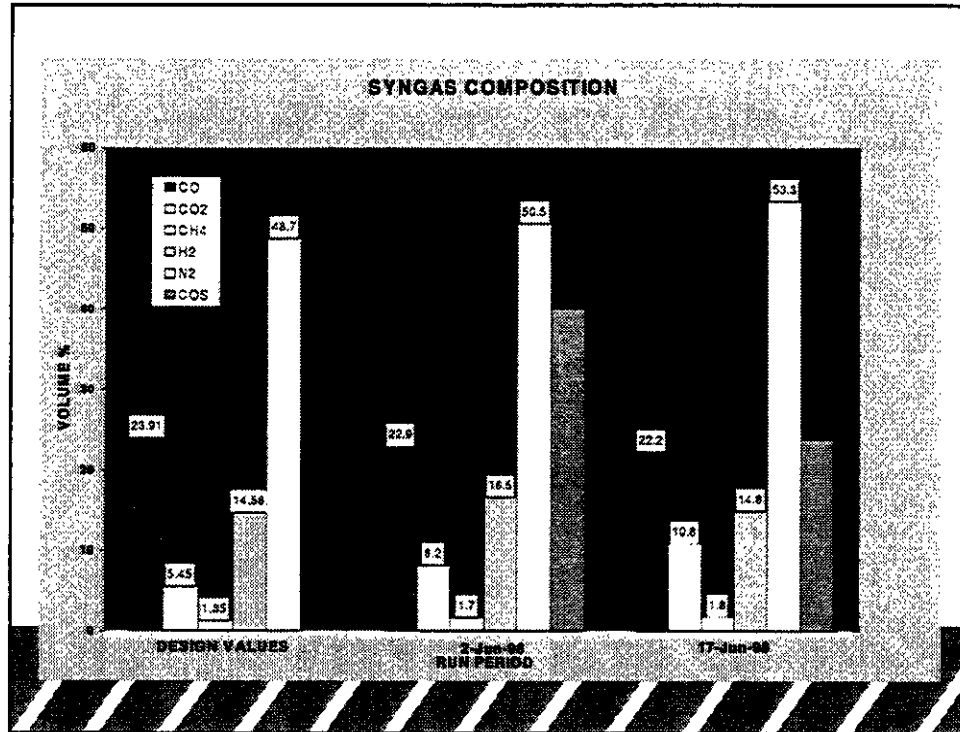
- Gasifier had only 10 successful runs
- longest In June, 1998 (3 runs) and February, 1999
- Averaged approximately 7 hours each, longest 12 hrs.
- Numerous Technical Problems Encountered and Resolved. Significant design issues not yet solved.

✓ Entire Facility Placed in Service in June '98 for Tax and Regulatory Purposes

1999 Performance Goals

- ◆ Combined Cycle: Achieve 90% availability
- ◆ Gasifier:
 - Achieve stable, sustained production of Syngas
 - Demonstrate sustained operation on Syngas
 - Successfully run gas turbine on Syngas



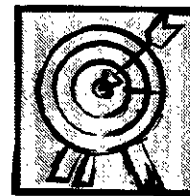


Operational Difficulties Fixed

◆ Significant Operational Problems Resolved:

– Gasifier:

- ◆ Nitrogen plant reliability
- ◆ Bin overpressurization
- ◆ Main coal feed tube
- ◆ Syngas flare
- ◆ Compressor issues
- ◆ Screw coolers, feeders, and lockhoppers



– Combined Cycle

- ◆ Turbine Generator coupling replaced (warranty fix)
- ◆ 2nd stage buckets replaced (warranty fix)

1998 Gasifier Startup Problems/Resolutions

◆ Additional Gasification Problems Resolved

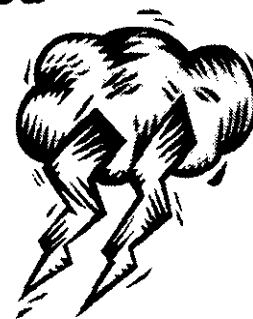
- Z-Sorb Failure
- Gasifier Slag Formation
- Gasifier Refractory Spalling
- Candle Breakage
- Candle Fail Safes
- Extraction Air Line Failure
- Coal Pile Fire In Dome

– Secondary Air Supply Failure

Design & Operational Difficulties Not yet Resolved

◆ Gasifier

- Gasifier annulus overtemperature
- Sulfator operability
- Fines transfer system
- Fines combustor
- HRSG / baghouse



◆ Combined Cycle

- Design fixes from GE to address coupling failure, 2nd stage shroud lifting problem

◆ Other issues may arise subsequently

Why has this been so hard?

- ◆ Very high degree of new technology
- ◆ High scaleup factors (to 4000%) on many components
- ◆ Design & Engineering deficiencies - some shouldn't have happened

Technology Outlook Questions

- ◆ Key Technology Questions in Decision Making are:
 - “Can Pinon be made to work acceptably--at 65% capacity factor or better?
 - “If So, at what Price?”
“Schedule?”
 - “What is the Outlook for the n'th plant?” (cost & performance)



Commercialization Issues & Outlook

- ◆ Numerous design and technology issues have been identified & fixed
- ◆ Some known problems being fixed at present
- ◆ No fundamental problems are currently known that will preclude successful demonstration, & subsequent commercialization
- ◆ Air-blown KRW technology continues to look viable, and can offer efficiency



COMMERCIAL-SCALE DEMONSTRATION OF THE LIQUID PHASE METHANOL (LPMEOH™) PROCESS: OPERATING EXPERIENCE UPDATE

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ABSTRACT

The Liquid Phase Methanol (LPMEOH™) process uses a slurry bubble column reactor to convert synthesis gas (syngas), primarily a mixture of carbon monoxide and hydrogen, to methanol. Because of its superior heat management, the process can utilize directly the carbon monoxide (CO)-rich syngas characteristic of the gasification of coal, petroleum coke, residual oil, wastes, or other hydrocarbon feedstocks. When added to an integrated gasification combined cycle (IGCC) power plant, the LPMEOH™ process converts a portion of the CO-rich syngas produced by the gasifier to methanol, and the unconverted gas is used to fuel the gas turbine combined-cycle power plant. In addition, the LPMEOH™ process has the flexibility to operate in a daily load-following pattern, coproducing methanol during periods of low electricity demand, and idling during peak times. Coproduction of power and methanol via IGCC and the LPMEOH™ process provides opportunities for energy storage for electrical demand peak shaving, clean fuel for export, and/or chemical methanol sales.

Since start-up in April 1997, performance of the LPMEOH™ Process Demonstration Plant has exceeded expectations. Following commissioning and shakedown activities, the first production of methanol from the plant occurred on April 2, 1997. Nameplate capacity of 260 tons per day (TPD) was reached for the first time on April 6, 1997, and production rates of over 300 TPD of methanol have been achieved. Several key milestones were achieved during 1998, including an

Research sponsored by the U.S. Department of Energy's Federal Energy Technology Center, under contract DE-FC22-92PC90543 with Air Products Liquid Phase Conversion Company, L.P., 7201 Hamilton Blvd., Allentown, PA 18195; fax: (610) 706-7299.

availability of over 99%. Since startup, over 35 million gallons of methanol have been produced at the LPMEOH™ Process Demonstration Plant, and plant availability has exceeded 96%.

This paper provides a description of the LPMEOH™ process, the commercial applications for the technology, and an update of the current plant performance results at the Kingsport site.

I. INTRODUCTION

The LPMEOH™ technology was developed during the 1980's with the financial support of the United States Department of Energy (DOE). The concept was proven in over 7,400 hours of test operation in a DOE-owned, 10 tons-per-day (TPD) Process Development Unit (PDU) located at LaPorte, Texas.¹ The first commercial-scale demonstration plant for the technology was sited at Eastman Chemical Company's (Eastman's) coal gasification facility in Kingsport, Tennessee, with the help of a \$92.7 million award under the DOE's Clean Coal Technology Program. Construction began in October of 1995 and concluded in January of 1997. After commissioning and startup activities were completed, operation began in April of 1997. During a four-year operating program, the LPMEOH™ Process Demonstration Plant will demonstrate the production of at least 260 TPD of methanol, and will simulate operation for the integrated gasification combined cycle (IGCC) coproduction of power and methanol application. The test plan will also seek to establish commercial acceptance of the technology and verify the fitness of the methanol product through a series of off-site, product-use tests. Total cost of the project, including the four-year demonstration test program, is forecast at \$213.7 million.

Air Products and Chemicals, Inc. (Air Products) and Eastman formed the "Air Products Liquid Phase Conversion Co., L.P." partnership to execute the project and own the LPMEOH™ Demonstration Plant. Air Products manages the overall program and provides technology analysis and direction for the demonstration. Air Products also provided the design, procurement, and construction of the plant (i.e., a turnkey facility). Eastman provides the host site, acquired the necessary permits, operates the demonstration plant, supplies the supporting auxiliaries and the synthesis gas (syngas), and takes the product methanol. Most of the product methanol is refined to chemical-grade quality (99.85 wt% purity) via distillation and used by Eastman as chemical feedstock elsewhere in their commercial facility. A portion of the product methanol has been withdrawn prior to purification (about 98 wt% purity) for use in off-site, product-use tests.

This paper reviews: the **Commercial Application** for the LPMEOH™ process technology; the **Demonstration Plant - Test Plans**, highlighting the operational plans to confirm the commercial application; and, the **Demonstration Plant - Current Performance Results**, highlighting the operating results achieved during the second year of operation.

II. COMMERCIAL APPLICATION

Technology Description

The heart of the LPMEOH™ process is the slurry bubble column reactor (Figure 1).

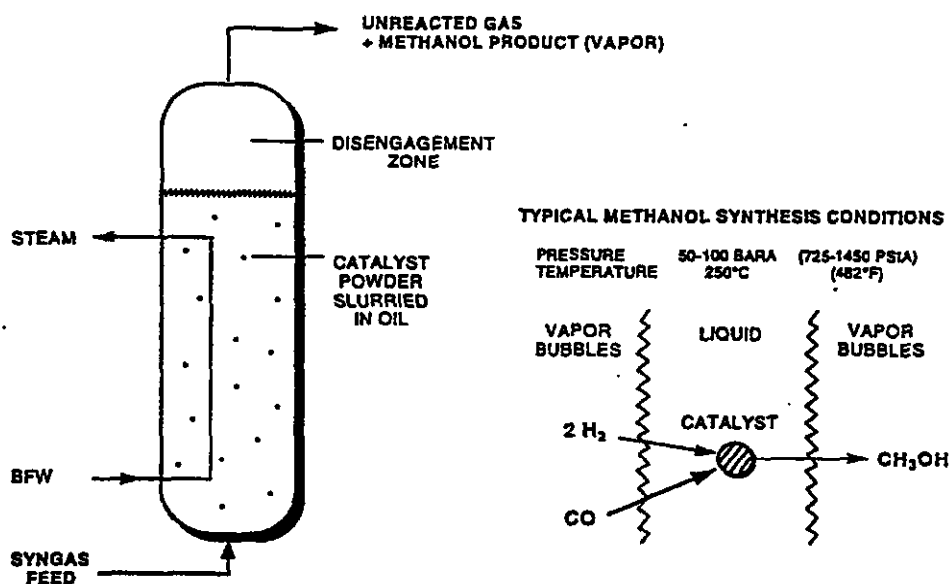


Figure 1. LPMEOH™ Reactor and Reaction Schematics

Conventional methanol reactors use fixed beds of catalyst pellets and operate in the gas phase. The LPMEOH™ reactor uses catalyst in powder form, slurried in an inert mineral oil. The mineral oil acts as a temperature moderator and a heat removal medium, transferring the heat of reaction from the catalyst surface via the liquid slurry to boiling water in an internal tubular heat exchanger. Since the heat transfer coefficient on the slurry side of the exchanger is relatively large, the heat exchanger occupies only a small fraction of the cross-sectional area of the reactor. As a result of this capability to remove heat and maintain a constant, highly uniform temperature through the entire length of the reactor, the slurry reactor can achieve much higher syngas conversion per pass than its gas-phase counterparts.

Furthermore, because of the LPMEOH™ reactor's unique temperature control capabilities, it can *directly* process syngas that is rich in carbon oxides (carbon monoxide and carbon dioxide). Gas-phase methanol technology would require that similar feedstocks undergo stoichiometric adjustment by the water gas shift reaction, to increase the hydrogen content, and subsequent carbon dioxide (CO₂) removal. In a gas-phase reactor, temperature moderation is achieved by recycling large quantities of hydrogen (H₂)-rich gas, utilizing the higher heat capacity of H₂, as compared to carbon monoxide (CO). Typically, a gas-phase process is limited to about 16% CO in the reactor inlet, as a means of constraining the conversion per pass to avoid excess heating. In contrast, for the LPMEOH™ reactor, CO concentrations in excess of 50% have been tested in the laboratory, at the PDU in LaPorte, and at the LPMEOH™ process demonstration plant without any adverse effect on catalyst activity.

A second distinctive feature of the LPMEOH™ reactor is its robust character and flexibility. The slurry reactor is suitable for rapid ramping, idling, and even extreme stop/start actions. The thermal moderation provided by the liquid inventory in the reactor acts to buffer sharp transient operations that would not normally be tolerable in a gas-phase methanol synthesis reactor. This characteristic is especially advantageous in the environment of electricity demand load-following in IGCC facilities.

A third differentiating feature of the LPMEOH™ process is that a high quality (generally greater than 97% purity) methanol product is produced directly from syngas rich in carbon oxides. Gas-phase methanol synthesis, which must rely on H₂-rich syngas, yields a crude methanol product with 4% to 20% water by weight. The product from the LPMEOH™ process, using CO-rich syngas, typically contains only 1% water by weight. As a result, raw methanol coproduced in an IGCC facility would be suitable for many applications at a substantial savings in purification costs. The steam generated in the LPMEOH™ reactor is suitable for purification of the methanol product to a higher quality or for use in the IGCC power generation cycle.

Another unique feature of the LPMEOH™ process is the ability to withdraw spent catalyst slurry and add fresh catalyst on-line periodically. This facilitates uninterrupted operation and also allows perpetuation of high productivity in the reactor. Furthermore, choice of replacement rate permits optimization of reactor productivity versus catalyst replacement cost.

IGCC Coproduction Options

The LPMEOH™ process is a very effective technology for converting a portion of an IGCC electric power plant's coal-derived syngas to methanol², as depicted in Figure 2. The process has the flexibility to handle wide variations in syngas composition. It can be designed to operate in a continuous, baseload manner, converting syngas from oversized gasifiers or from a spare gasifier. Alternatively, the process can be designed to operate only during periods of off-peak electric power demand, consuming a portion of the excess syngas and reducing the electricity output from the combined-cycle power unit. In this scenario, the gasification unit continues to operate at full baseload capacity, so that the IGCC facility's major capital asset is always fully utilized.

In either baseload or cycling operation, partial conversion of between 20% and 33% of the IGCC plant's syngas is optimal, and conversion of up to 50% is feasible. The required degree of conversion of syngas, or the quantity of methanol relative to the power plant size, determines the design configuration for the LPMEOH™ plant. In its simplest configuration, syngas at maximum available pressure from the IGCC power plant's gasifier system passes once-through the LPMEOH™ plant and is partially converted to methanol without recycle, water-gas shift, or CO₂ removal. The unreacted gas is returned to the IGCC power plant's combustion turbines. If greater syngas conversion is required, different plant design options are available.³

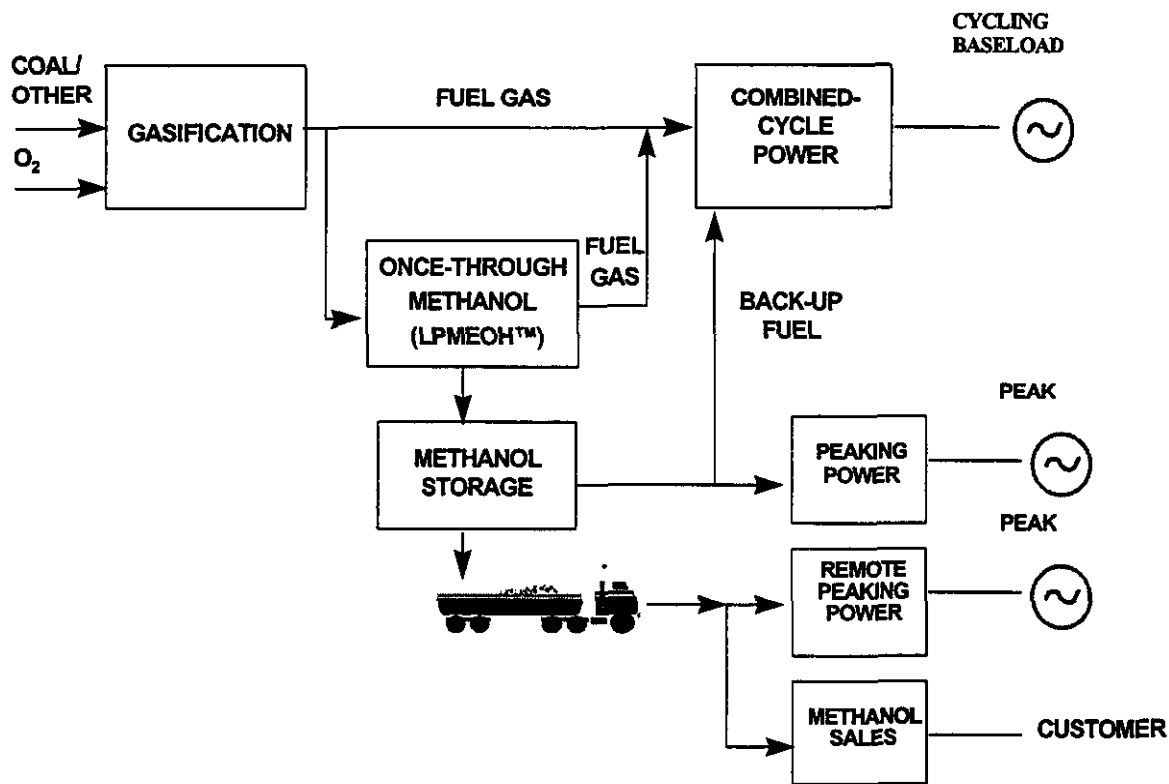


Figure 2.
Once-through Methanol Coproduction with IGCC Electric Power

Economics for Baseload Coproduction of Methanol and Power

Design studies for the LPMEOH™ process have focused principally on the aforementioned IGCC applications. A comparison of the cost of methanol as produced from the LPMEOH™ Process and from a conventional gas phase process as applied to a generic 500 TPD methanol plant as part of an IGCC coproduction facility was performed.⁸ The cost of methanol is calculated as the sum of three terms: the methanol conversion cost (which includes the fixed and operating costs for the methanol unit), the distillation cost to produce fuel grade methanol, and the syngas cost from the IGCC facility. A proprietary cost estimation screening program was used to calculate the methanol conversion cost and the distillation cost from the LPMEOH™ Process and the gas phase process for various syngas supply pressures and on-stream factors. Based on this analysis, the methanol conversion cost from the LPMEOH™ Process is \$0.02 to \$0.07 per gallon lower than from the gas phase methanol process depending on syngas supply pressure and composition and facility on-stream factor.

The LPMEOH™ Process can directly convert coal-derived syngas which is rich in CO, to produce a crude methanol product with nominally about 1 wt% water. Whereas, gas phase methanol synthesis results in a crude methanol product with 2-20 wt% water, depending on the amount of CO₂ in the syngas which is converted to methanol and water. This results in lower

purification cost for the LPMEOH™ process for the applications where high purity distillation is not required, such as Fuel Grade and MTBE Grade products.

Methanol coproduction, by IGCC and the once-through LPMEOH™ Process, does not require large methanol plant sizes to achieve good economies of scale. The gasification plant is necessarily at a large economical scale for power generation, so the syngas manufacturing economies are already achieved. Methanol storage and transport economies are also achieved by serving local markets, and realizing freight savings over competing methanol, which is usually shipped from the U. S. Gulf Coast.

III. DEMONSTRATION PLANT - TEST PLANS

The preceding Commercial Application section highlighted the advantages of the LPMEOH™ Process as part of an IGCC electric power generation system. To confirm these commercial advantages during operations, the demonstration test plan incorporates, but is not limited to, the following commercially important aspects of IGCC integration:

- Syngas compositions will vary with the type of gasification process technology and feedstock used in the power generation application. Therefore, operation over a wide variety of syngas compositions will be demonstrated.
- Catalyst life, operating on coal-derived syngas, must be demonstrated over a long period of time. Major parameters include reactor operating temperature, concentration of poisons in the reactor feed gas, and catalyst aging and attrition.
- Reactor volumetric productivity must be optimized for future commercial designs. Parameters include: high inlet superficial velocity of feed gas, high slurry catalyst concentration, maximum expanded slurry level, and removal of the heat of reaction.
- Methanol Product, as produced by the LPMEOH™ reactor from syngas rich in carbon oxides, must be suitable for its intended uses. Off-site methanol product-use testing will confirm the product specifications needed for market acceptability.

Although generation of electric power is not a feature of the demonstration project at Kingsport, the demonstration test plan is structured to provide valuable data related to the following:

- coproduction of electric power and value-added liquid transportation fuels and/or chemical feedstocks from coal. This coproduction requires that the partial conversion of syngas to storable liquid products be demonstrated.
- energy load-following operations that allow conversion of off-peak energy, at attendant low value, into peak energy commanding a higher value. This load-following concept requires that on/off and syngas load-following capabilities be demonstrated.

Three key results will be used to judge the success of the LPMEOH™ Process demonstration during the four years of operational testing:

- resolution of technical issues involved with scaleup and first time demonstration for various commercial-scale operations;
- acquisition of sufficient engineering data for future commercial designs; and,
- industry or commercial acceptance.

The demonstration test plan provides flexibility to help meet these success criteria. Annual operating plans, with specific targeted test runs, will be prepared, and revised as necessary. These plans will be tailored to reflect past performance, as well as commercial needs. The LPMEOH™ operating test plan outline, by year, is summarized in Table 1.

The demonstration test plan encompasses the range of conditions and operating circumstances anticipated for methanol coproduction with electric power in an IGCC power plant. Since Kingsport does not have a combined-cycle power generation unit, the tests will simulate the IGCC application. In addition, the test program will emphasize test duration. The minimum duration for a test condition, apart from the rapid ramping tests, is 2 weeks. Numerous tests will have 3 to 6 week run periods, some 8 to 12 weeks, and a few key basic tests of 20 to 30 weeks.

The ultimate goal of the demonstration program is to reach a stable, optimized operating condition, with the best combination of the most aggressive operating parameters. These parameters, such as reactor superficial gas velocity, slurry concentration, and reactor level, will allow maximum reactor productivity to be achieved. Debottlenecking limitations of the demonstration plant will be an on-going goal during the demonstration program.

Table 1. LPMEOH™ Demonstration Test Plan Outline	
Accomplishments:	
<u>1997</u>	Catalyst Aging Catalyst Life vs. LaPorte process development unit and Lab Autoclaves Process Optimization / Maximum Reactor Productivity Catalyst Slurry Concentration (increasing to 40 wt%) Reactor Slurry Level Catalyst Slurry Addition Frequency Gas Superficial Velocity Long-term Continuous Test Period - 31 days Establishment of Baseline Condition
<u>1998</u>	Catalyst Attrition/Poisons/Activity/Aging Maximum Catalyst Slurry Concentration (exceeding 40 wt%) Alternative Catalyst Long-term Continuous Test Periods - 65 and 94 days 99.7 % Availability
Future Work:	
<u>1999 - 2001</u>	Catalyst Slurry Addition and Withdrawal at Baseline Condition Tests Continued Catalyst Attrition/Poisons/Activity/Aging Simulation of IGCC Coproduction for: <ol style="list-style-type: none"> 1. Syngas Composition Studies for Commercial Gasifiers Texaco, Shell, Destec, British Gas/Lurgi, Other Gasifiers 2. IGCC Electrical Demand Load-Following: Rapid Ramping, Stop/Start (Hot and Cold Standby). 3. Additional Industry User Tests Maximum Throughput/Production Rate Temperature Programming In-Situ Catalyst Activation (under evaluation) Stable, extended Operation at Optimum Conditions Potential Liquid-Phase Dimethyl Ether (LPDME™) Test

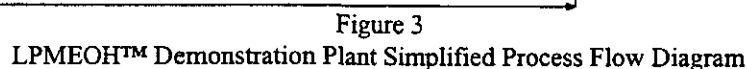
IV. DEMONSTRATION PLANT - PERFORMANCE RESULTS

Kingsport Site

Eastman began coal gasification operations at Kingsport, TN in 1983. Texaco gasification converts about 1,000 tons-per-day of high-sulfur, Eastern bituminous coal to syngas for the

Because the gasification facility produces individual streams of clean balanced syngas (Balanced Gas), CO (CO Gas), and H₂-rich gas (H₂ Gas), the LPMEOH™ Demonstration Plant design includes the capability to blend these streams into a wide range of syngas compositions. This flexibility enables the plant to simulate the feed gas composition available from any commercial gasifier.

Figure 3 shows a simplified process flow diagram of the LPMEOH™ Demonstration Plant. Approximately half of the Balanced Gas fresh feed to the existing methanol unit is diverted to the LPMEOH™ Demonstration Plant, where it combines with the high-purity CO Gas and passes through an activated carbon guard bed. This bed removes iron and nickel carbonyls, which are poisons to methanol synthesis catalyst, down to ppb levels. The third feed stream, H₂ Gas, is the hydrogen-rich purge exiting the existing methanol unit. Since the H₂ Gas is at lower pressure than the other two feed streams, it is combined with the Recycle Gas stream, made up of unconverted syngas from the LPMEOH™ reactor, and compressed in the recycle compressor.



These two pairs of streams are then combined to form a single high pressure reactor feed gas stream that is preheated in the feed/product economizer against the reactor effluent. The feed gas is then sparged into the LPMEOH™ reactor, where it mixes with the catalyst slurry and is partially converted to methanol vapor, releasing the heat of reaction to the slurry. The slurry temperature is controlled by varying the steam temperature within the heat exchanger tubes, which is accomplished by adjusting the steam pressure.

Disengagement of the effluent gas (methanol vapor and unreacted syngas) from the catalyst/oil slurry occurs in the freeboard region of the reactor. Any entrained slurry droplets leaving the top of the reactor are collected in the cyclone separator. The product gas passes through the tubeside of the feed/product economizer, where it is cooled against the reactor inlet gas stream. Any condensed oil droplets are collected in the high-pressure oil separator and then returned to the reactor with the entrained slurry from the cyclone separator.

The product gas is cooled further in a series of air-cooled and cooling water exchangers, whereupon the product methanol condenses and collects in the high pressure methanol separator. Most of the unreacted syngas returns to the reactor after undergoing compression in the recycle compressor. The balance of the unreacted syngas is purged to the Eastman fuel gas system.

The condensed methanol contains dissolved gases, water, trace oil, and some higher alcohols. These impurities are removed in a two-column distillation train that produces a methyl acetate-grade methanol feed product. The bottom draw from the second column is a crude methanol stream heavy in higher alcohols, water, and any oil carried over from the reactor. This stream is sent to the existing distillation system for recovery of the methanol and disposal of the byproducts. Stabilized, fuel-grade methanol for off-site product-use testing will be produced at limited times during the demonstration period by using only the first distillation column.

Catalyst slurry is activated in the catalyst reduction vessel, which is equipped with a heating/cooling jacket, utility oil skid, and agitator. Pure CO, diluted in nitrogen, acts as the reducing agent. During the activation procedure, slurry temperature is carefully increased while monitoring consumption of CO to determine when the catalyst is completely reduced. At the end of this procedure, the catalyst is fully active and can be pumped directly to the reactor. Catalyst inventory is maintained by a combination of catalyst addition and withdrawal: as fresh catalyst slurry is added to the LPMEOH™ reactor, catalyst inventory is maintained by withdrawing an equivalent amount of partially deactivated or spent slurry.

Initial Operation

Table 2 summarizes the commissioning and startup milestones at the LPMEOH™ Demonstration Plant.

Table 2.

LPMEOH™ Demonstration Plant Milestones

• Groundbreaking	October 1995
• Plant Mechanically Complete	January 1997
• Eastman Begins Commissioning	February 1997
• Completed Startup	April 1997
• Achieved Design Catalyst Life	February 1998
• Achieved Design Catalyst Loading	September 1998
• Successful Reactor Inspection	March 1999
• Availability in 1998	99.7 %

After activation of nine 1-ton batches of methanol synthesis catalyst, the reduced catalyst slurry was pressure-transferred from a maintenance tank to the LPMEOH™ reactor on April 1, 1997. In less than two weeks of operation, the LPMEOH™ Demonstration Plant met several of its short-term performance goals. Methanol production reached the nameplate capacity of 260 TPD, and a stable test period at over 300 TPD of methanol revealed no system limitations, either in the reactor or distillation areas. The rapid progression from first introduction of syngas to stable operation at greater than nameplate capacity is an indication of the robust nature of the LPMEOH™ process. The startup also proceeded without injury or environmental incidents.

Since initial operation in April 1997 to the end of calendar year 1998, the LPMEOH™ Demonstration Plant achieved an availability in excess of 96%. During the first year of operation, a 31-day continuous run was achieved which helped to illustrate and confirm the overall system reliability. The H₂/CO ratio in the reactor feed stream was varied from 0.4 to 5.6 with no negative effects on performance. The results pertaining to gas holdup (the volume fraction of the reactor occupied by gas), an important design parameter for slurry reactors, have provided initial confirmation of the equipment scale-up parameters for the LPMEOH™ reactor. Important parameters such as high inlet superficial velocity of reactor feed gas, maximum expanded slurry level, and the overall heat transfer coefficient of the internal heat exchanger have been demonstrated at 115 - 120% of design levels.

Initial catalyst life data indicated an accelerated change in performance occurred; whereas, the remaining operation from June 1997 through November 1997 matched the typical activity loss measured in the laboratory. Figure 4 shows performance results from the LPMEOH™ reactor during the first several months of operation. The data are reduced to a ratio of rate constant pre-exponential factors (actual vs. design value for fresh catalyst), using an in-house kinetic model, to eliminate the effects of changing feed composition or operating conditions. Typical

4th Annual Clean Coal Technology Conference Proceedings



exponential decay will appear as a straight line on a log-plot, as shown. The curve fit to data from a 4-month test at the LaPorte PDU in 1988/89 and laboratory autoclave data from 1996 are included for reference.

An important feature of the LPMEOH™ Process is the ability to remove spent catalyst from the reactor during operation; this also affords the opportunity to examine samples for changes in the microscopic structure and/or chemical make-up of the catalyst with time. Analyses of early samples from Kingsport have indicated a step-change increase in the concentration of iron on the catalyst surface during the initial six weeks which cannot be correlated to the presence of iron carbonyl in the feed gas streams. This finding is most likely related to the detection of post-construction debris within various parts of the facility. Higher than expected levels of arsenic were also found on the catalyst samples. After an initial operating period of seven months, the reactor was drained and another partial charge of fresh catalyst was activated during December of 1997.

Kingsport LPMEOH™ Catalyst Life (First Campaign: 1997)

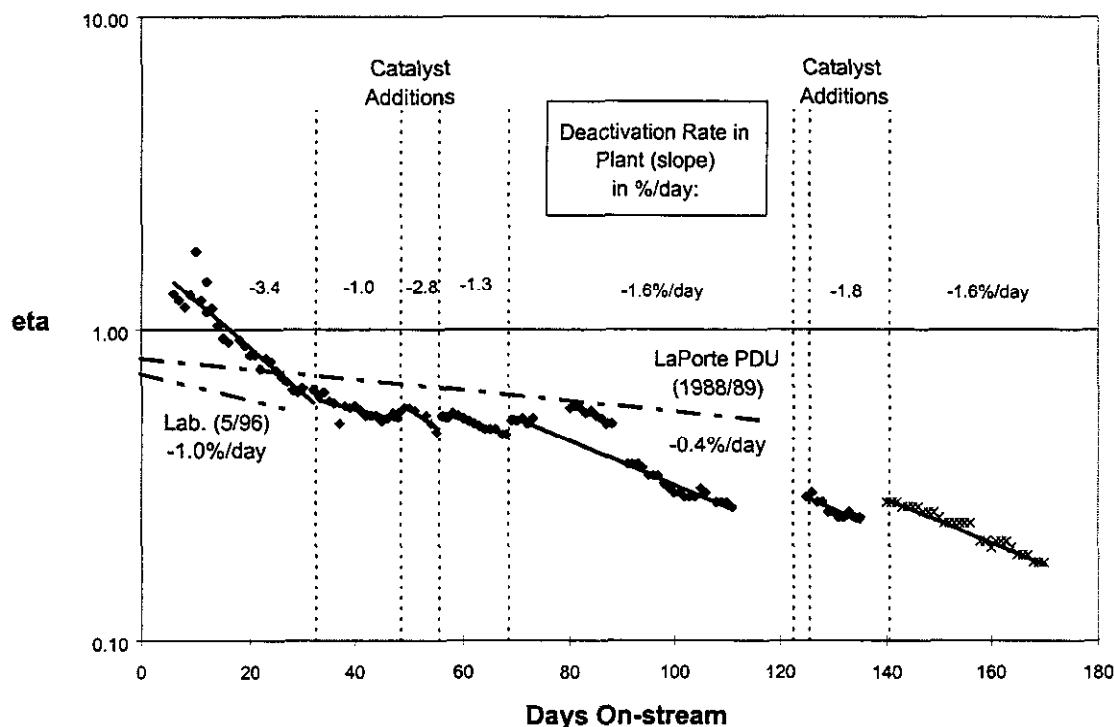


Figure 4
Catalyst Life During Campaign 1

Performance Results - December 1997 to December 1998

Catalyst life was exceedingly improved during the second campaign of operation which began in December 1997. Iron levels in the catalyst samples did not increase significantly during this operating period. This further supports the conclusion that early elevated iron levels were a function of construction debris or start-up conditions. During this second operating campaign, the reactor temperature was lowered to 235°C, a reduction from the 250°C temperature used during initial operation. This was an effort to further investigate catalyst performance issues. The calculated catalyst activity curve since the restart is included in Figure 5. The combined results of lowered reactor temperature and reduced poison deposition were a catalyst life or deactivation rate that met or exceeded the life achieved in the LaPorte PDU during several long duration operation periods. A parallel lab test trial conducted on plant gas is shown for comparative purposes.

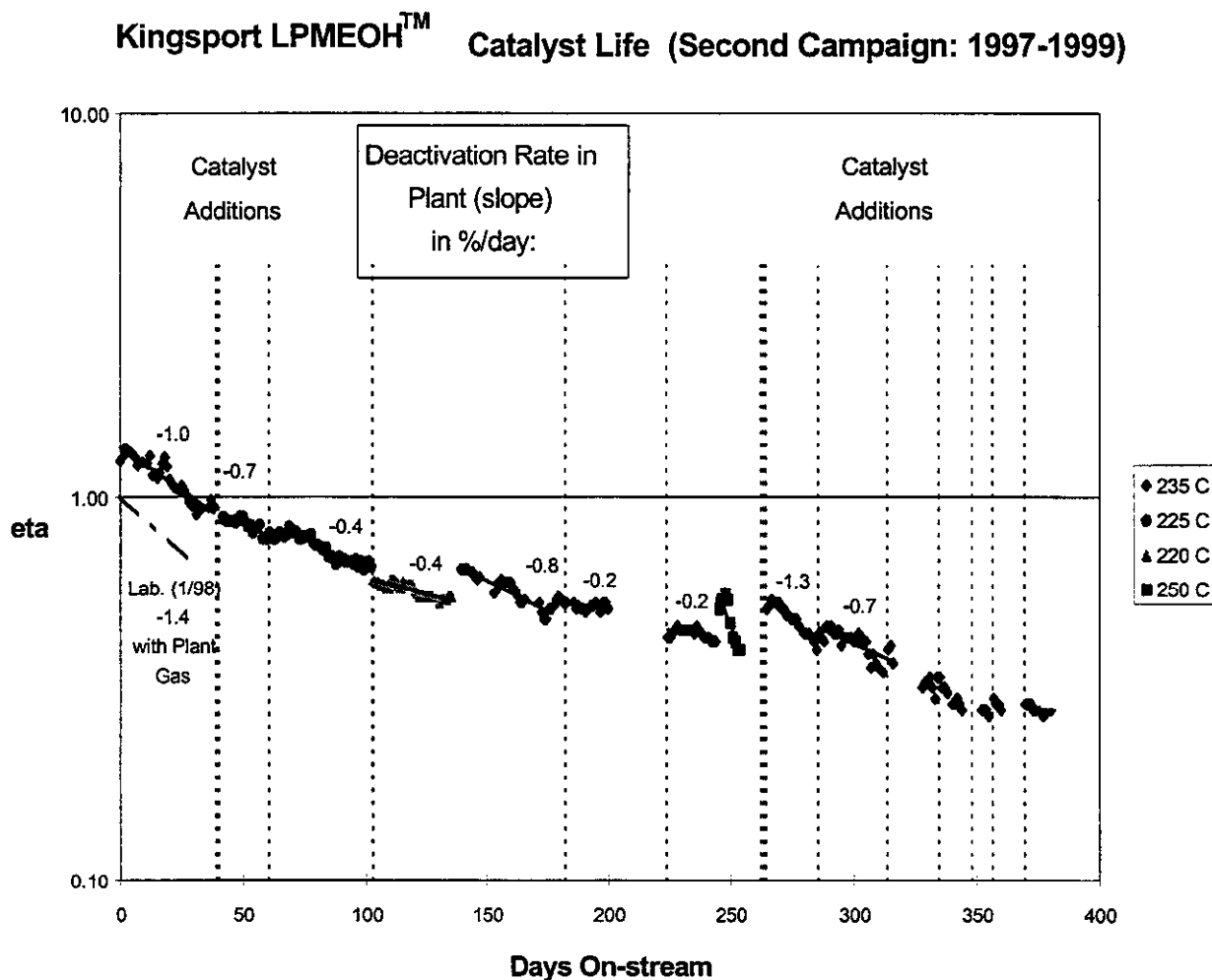


Figure 5
Catalyst Life During Campaign 2

Although the presence of iron on the catalyst samples was not significant during this second theater of operations, levels of arsenic and, to a lesser degree, sulfur increased noticeably. The arsenic and sulfur compounds are potential catalyst poisons found in the coal-derived syngas. However, there is no definitive correlation linking either species to catalyst deactivation results. To further mitigate the effects of arsenic on the catalyst, arsenic removal capacity will be increased in guard beds located both upstream and within the LPMEOH™ Process. The level of all potential poisons will continue to be monitored during the on-going plant operation.

A major area of consideration has been to study the effect of higher catalyst loading on reactor performance. During the latter part of 1998, a concerted effort was made to increase the catalyst loading and slurry concentration through judicious additions of catalyst batches. Figure 6 shows the concentration of the reactor slurry during this second campaign of operation. As shown in the figure, the design catalyst concentration of 40% was significantly exceeded with values as high as 49% experienced. In addition, a catalyst inventory loading of over 150% of design was achieved in the reactor during this period of operation. No negative effects, including mass transfer limitations, were experienced during this operating period.

A key design feature of the slurry reactor is its ability to manage the heat release from exothermic reactions such as methanol synthesis. Since startup, the heat management performance of the LPMEOH™ reactor has been outstanding. In fact, the temperature difference between the process and the steam system is essentially constant along the length of the reactor. The maximum temperature profile in the slurry, as measured by 35 thermocouples at various axial and radial intervals, is 3 to 4°C axially, and less than 1°C radially. The heat transfer coefficient for the internal heat exchanger has exceeded the design value and has indicated little fouling to date. The absence of slurry-side fouling has been additionally confirmed by inspection of the heat exchanger during a scheduled outage.

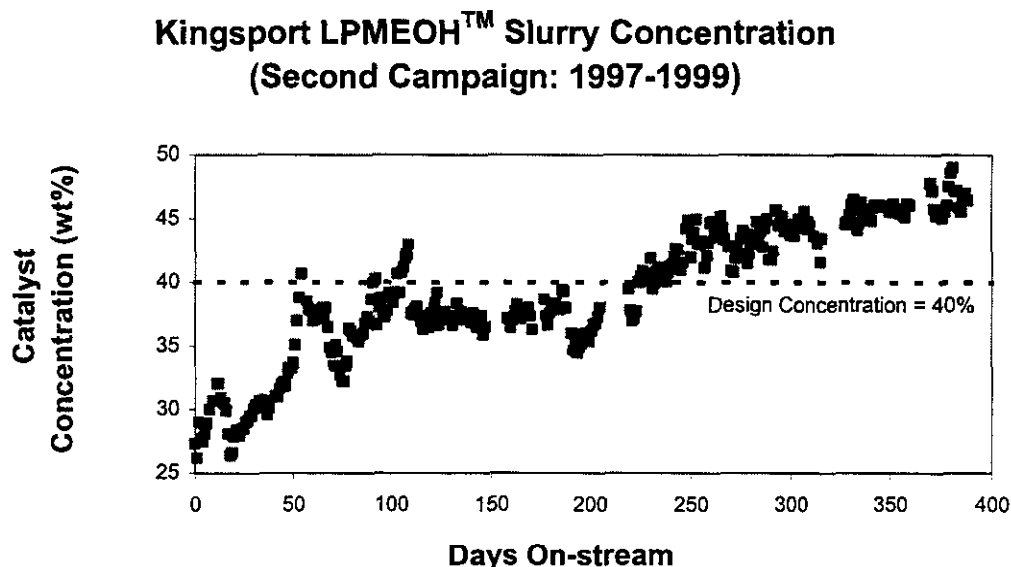


Figure 6
Catalyst Slurry Concentration During Campaign 2

Gas holdup, the volume fraction of the reactor occupied by gas, is an important parameter for sizing a LPMEOH™ reactor. A low holdup may indicate poor mixing, likely resulting in sub-optimum reactor performance. A high gas holdup leaves less space in the reactor for catalyst inventory. Pressure-differential transmitters, at regular intervals along the length of the reactor, measure the average density in discrete volumes of slurry. The gas holdup is calculated for each section using the known catalyst loading and the physical properties of the three phases. The results pertaining to gas holdup and heat management have provided initial confirmation of the equipment scale-up parameters for the LPMEOH™ reactor; additional operating time will verify long-term performance and the impact of changes in other operating conditions, such as gas composition and flowrate.

The LPMEOH™ Demonstration Plant was shutdown in early March 1999 to begin a scheduled bi-annual inspection of all pressure vessels as required by Tennessee state code. Catalyst slurry was pressure transferred from the LPMEOH™ reactor to the slurry holding tank and contained during the outage period. No issues were observed with any of the units evaluated as part of the code inspection. The walls and internal heat exchanger of the LPMEOH™ reactor internally showed no evidence of erosion, pitting, or fouling by catalyst slurry. The inspection activities were completed in a timely manner and the catalyst slurry was transferred back from the slurry holding tank to the LPMEOH™ reactor on 13 March 1999 for a restart of operation. Following the inspections, the LPMEOH™ Demonstration Plant was re-started on 14 March 1999 with the transferred catalyst inventory.

A scheduled complex-wide outage followed in mid-April at the Eastman Chemical facility as the LPMEOH™ Demonstration Plant passed into its third year of operation. During this outage, the LPMEOH™ Demonstration Plant was held with a full catalyst load under stand-by conditions for over 11 days. The successful re-start after this outage further speaks to the robustness of the technology to handle start-up, shutdown, and stand-by conditions.

Future Activities

During 1999, efforts will continue to sample the catalyst from the reactor and monitor plant performance to quantify the long-term catalyst aging characteristics under coal-derived syngas. In addition, operations with CO-rich syngas and other reactor feed gas compositions are planned.

V. CONCLUSION

The LPMEOH™ Process is now being demonstrated at commercial scale under the DOE Clean Coal Technology Program. The demonstration plant, located at Eastman Chemical Company's Kingsport, Tennessee coal gasification facility, has produced in excess of the 260 TPD of methanol nameplate capacity from coal-derived syngas. Since startup of the unit in April of 1997, overall availability has exceeded 96 %, while the more recent campaign in calendar year 1998 achieved 99.7 % availability. The startup and initial operation proceeded without injury or

environmental incidents, and Eastman has accepted all of the greater than 35 million gallons of methanol produced at the LPMEOH™ Demonstration Plant for use in downstream chemical processes.

Successful demonstration of the LPMEOH™ technology will add significant flexibility and dispatch benefits to IGCC electric power plants, which traditionally have been viewed as strictly a baseload power generation technology. Now, central clean coal technology processing plants, making coproducts of electricity and methanol, can meet the needs of local communities for dispersed power and transportation fuel. The LPMEOH™ Process provides competitive methanol economics at small methanol plant sizes, and a freight and cost advantage in local markets vis-à-vis large offshore remote gas methanol. Methanol coproduction studies show that methanol can be produced at economically competitive levels from an abundant, non-inflationary local fuel source, such as coal. The coproduced methanol may be an economical hydrogen source for small fuel cells, and an environmentally advantaged fuel for dispersed electric power.

VI. ACKNOWLEDGMENT

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TECHNICAL SESSION II

Combustion Systems: How Can CCTs
Meet the Needs, Part II

HEALY CLEAN COAL PROJECT: 1998 COMBUSTOR AND SPRAY DRYER ABSORBER CHARACTERIZATION TESTING

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ABSTRACT

The Healy Clean Coal Project (HCCP), selected by the U.S. Department of Energy under Round III of the Clean Coal Technology Program, is currently in its test operations phase. After more than five years of planning, design engineering and permitting activities, the project celebrated its ground-breaking ceremony at Healy, Alaska on May 30, 1995. Most of the major plant equipment was delivered to the Healy site 250 miles north of Anchorage, Alaska in 1996. Construction of the plant was completed in November 1997, with coal-fired operations starting in January 1998. The project status, its participants, description of the HCCP technology, and the 1998 operational performance of the combustion system and flue gas clean-up system are presented in this paper.

The TRW Clean Coal Combustion System is designed to minimize NO_x emissions, achieve very high carbon burnout, and remove the majority of flyash from the flue gas prior to the boiler. The TRW system also provides the first step of a three-step process for controlling SO₂ by converting limestone to flash calcined lime that subsequently absorbs SO₂ within the boiler. The majority of SO₂ is removed downstream of the boiler, using Babcock & Wilcox's (B&W's) activated spray-dryer absorber (SDA) system, which utilizes the flash calcined material (flash calcined lime + flyash) produced by the TRW system. Since most of the coal ash is removed by the combustors, the flash calcined material is rich enough in calcium content such that the SDA can be operated solely on recycled lime, eliminating the need to purchase or manufacture lime for the backend scrubbing system.

HCCP is the first utility-scale demonstration of the TRW clean coal combustion technology. During 1998, approximately 5,000 hours of plant thermal operation were accumulated, with approximately 4,500 hours of coal-fired operating time. Both run-of-mine (ROM) and ROM/Waste coal blends were tested in the combustion system. For the majority of the testing, the coal combustors were operated in conjunction with B&W's activated recycle spray dryer absorber system. To date testing has shown:

- ◆ *ability to achieve low NO_x emissions simultaneously with low CO emissions and high carbon burnout*
- ◆ *good combustion efficiency and high slag removal prior to the furnace*

- ◆ good limestone calcination efficiency
- ◆ consistent achievement of SO_2 emissions less than 0.10 lb / MMBtu.

This paper presents the results of coal-fired test operations from June 12 through December 21, 1998. During this period of time, approximately 3300 hours of plant thermal operation was accumulated, with approximately 3200 hours of coal-fired operation. The majority of test operations were at full load, 50 MW_e. The emission data presented includes all coal-fired operations during this period of time, including, in most cases, start-up and shutdown operations. Not included herein is emission data during: 1) January through June 11, 1998, which primarily consisted of coal-firing start-up and shake down activities and was prior to certification of the Continuous Emissions Monitoring System (CEMS) and, 2) oil-fired only operation. The emission levels of NO_x , CO, and SO_2 , were lower than permitted emission limits. From June through December 1998, the demonstrated environmental performance while burning ROM or ROM/Waste Coal Blends was as follows:

<i>NO_x Emissions:</i>	<i>0.208 to 0.278 lb NO_x / MMBtu (0.245 average)</i>
<i>SO₂ Emissions:</i>	<i>0.01 to 0.09 lb SO₂ / MMBtu (0.036 average)</i>
<i>Ca/S Ratio:</i>	<i>1.0 to 3.0 (typically less than 2.0)</i>
<i>CO Emissions:</i>	<i>0.01 to 0.13 lb CO / MMBtu (0.038 average)</i>
<i>Ash Removal:</i>	<i>80 to 90% (including less than 5% bottom ash)</i>

If published data from the Continuous Emission Monitoring system is used, which includes oil-fired only data during start-up and shutdown, the average NO_x emissions over this time period is 0.25 lb NO_x / MMBtu versus the 0.245 average shown in the table above. The NO_x emission levels presented above were achieved prior to any optimization of furnace air staging or furnace O₂ levels. In general, the lowest NO_x emission levels at full load were achieved at lower furnace O₂ levels (3.0-3.5%), without any significant increase in plant CO emissions. The CO emissions were measured by a CO analyzer located at the furnace exit. This analyzer is not part of the Continuous Emission Monitoring System.

Demonstration Test operations are continuing during 1999. The focus of the 90-day Demonstration Testing planned for 1999 will be to accumulate data on availability and sustained operations. On-going preparations for long duration operation include addressing some previously identified plant operational and/or hardware durability problems, and includes such items as improving mill exhaust fan erosion resistance, mitigating slag/ash falls from the furnace hopper slope, and selecting optimal flame scanner locations. In addition, fine-tuning of the control system will be performed in order to improve system response time to load changes.

The \$242 Million project is owned and financed by the Alaska Industrial Development and Export Authority (AIDEA), and is cofunded 48% by the U.S. Department of Energy (DOE). Golden Valley Electric Association of Fairbanks, Alaska will be the contract operator and provided the plant operators for the testing. Usibelli Coal Mine of Healy Alaska provided the Run-of-Mine and Waste coals fired during the combustor characterization testing.

I. BACKGROUND

The multistage coal combustion technology demonstrated at the Healy Clean Coal Project (HCCP) power plant started at TRW with Low-NO_x utility oil burners in the 1970s and with pressurized magnetohydrodynamic (MHD) coal combustors in the early 1980s. Initial tests at TRW of an atmospheric pressure coal combustor at 10 MMBtu/hr in 1982 were followed by testing of a 40 MMBtu/hr industrial size combustor using a wide variety of coals to obtain extensive data on combustion, slag removal, NO_x and SO₂ emission and particulate carry over. A retrofit demonstration at a Cleveland, Ohio manufacturing plant was started in 1984 and over 10,000 hours of operation were accumulated while providing plant steam at high availability. Fifteen different coals with a wide range of physical properties were tested in this industrial-size coal combustor:

Moisture	1.36 % to 31.7 %
Ash	4.39 % to 27.32 %
Volatiles (dry, ash free)	10.6 % to 60.8 %
Nitrogen (dry, ash free)	0.95 % to 1.9 %
Sulfur (dry, ash free)	0.48 % to 4.59 %
Higher Heating Value (HHV)	7,358 Btu/lb to 13,061 Btu/lb
Ash Fusion Temp. (T-250)	2,118 deg F to 2,900 deg F

During the early 1990s, a utility-scale prototype version of the Healy precombustor and a 7.5 ton/hour direct coal feed system were successfully tested at TRW's Fossil Energy Test Site as part of the Healy Clean Coal Project. More than five years of planning and permitting culminated in spring 1995 with the start of construction on the 50 MW_e (net) HCCP power generation unit. During the summer of 1995, earthwork, foundation and structural steel work began with construction and erecting of all equipment continuing through late 1997. Construction was completed in November 1997, with coal-fired operations starting in January 1998.

II. TECHNOLOGY

The Healy Clean Coal Project integrates a slagging, multi-staged coal combustor system with an innovative sorbent injection/spray dryer absorber/baghouse exhaust gas scrubbing system. Twin 350 MMBtu/lb combustors designed by TRW are used to supply hot gases to a conventional Foster Wheeler bottom-fired boiler. The flue gas cleaning equipment was supplied by Babcock & Wilcox (B&W) based on technology developed by Joy Environmental Technologies of Houston, Texas and NIRO Atomizer of Denmark.

The first step in the Healy Clean Coal Project combustion process, shown in schematic format in Figure 1, is the pulverized coal feed system which consists of coal silos, Foster Wheeler MBF 21.5 coal pulverizers and exhauster fans, and the TRW coal feed system. The purpose of this system is to ensure a steady feed of coal (over a wide range of physical properties) to both combustion stages. The second step in this process is the TRW multi-staged coal combustor

which utilizes a multi-staged combustion process to minimize the formation of nitrogen oxides while burning a wide variety of coals including "hard to burn" coals. This combustor system melts and removes most of the coal mineral contaminants as slag. Pulverized limestone is injected prior to the combustor-boiler interface to provide for SO_2 removal from the combustion gases. The limestone is converted by heat in the combustion gases to flash calcinated material (high surface area lime + ash particles, called FCM) which reacts with the SO_2 in the combustion gases and removes the SO_2 as calcium sulfate. The unreacted FCM and sulfates are captured and recycled within the B&W spray-dryer absorber system downstream of the boiler to further reduce the SO_2 content in the combustion gases prior to the exhaust stack.

At HCCP, the two coal combustors are installed side by side and fire the boiler from the bottom upwards. An isometric view of the boiler and combustion system used in the HCCP is shown in Figure 2. Each combustor has its own dedicated coal storage, grinding, and feed system. Crushed coal is discharged from a storage silo into a pulverizer via a coal feeder/weighing scale. The pulverized coal and pulverizer sweep air (or primary air) is boosted in pressure to 60 inches of water (gauge) (1.15 atm) by the mill exhaust fan. This pressure is necessary to overcome the pressure drop through a non-storage coal feed/splitter subsystem that enables the coal to be split and fed into the precombustor and slagging stage. The coal feed/splitter subsystem also separates a major portion of the primary air and diverts this air to NO_x ports located in the boiler furnace. This helps in reducing the amount of cold air going into the combustor, thereby increasing the temperature of the combustion gases to promote slagging conditions over the entire range of coal ash melting temperatures.

Pulverized coal is fed to both the precombustor and slagging stages of the combustor. The precombustor portion of the coal is fed directly to a coal burner located in the headend of the precombustor. The slagging stage portion of the coal is split into six parts and injected into the head-end of the slagging stage via six injection ports. From the slagging stage, the combustion gases enter the slag recovery section where the gases are directed vertically upwards into the furnace through an interface opening in the sloping bottom of the furnace. The fuel rich combustor exhaust is intercepted by the boiler NO_x port air, where final air is added to complete combustion. Optional over-fire-air can also be introduced to provide further air staging for supplemental NO_x and temperature control.

A single limestone feed subsystem services both combustors. Pulverized limestone, stored in a silo, discharges via a weigh scale feeder to a rotary air-lock and a two-way splitter. A separate air-driven eductor is used at each leg of the splitter to transport the limestone-air mixture to a single limestone injector located on the side of the slag recovery section. The limestone particles flash-calcine to highly reactive lime with high surface area. These particles remove some of the SO_2 from the combustion gases as they pass through the furnace. The FCM particles are collected and utilized by the B&W spray-dryer absorber system to remove most of the remaining SO_2 in the combustion products, typically resulting in less than 10% of the sulfur contained in the coal exiting as SO_2 with the plant stack gases. Final FCM and fly ash particulate control is accomplished in the baghouse.

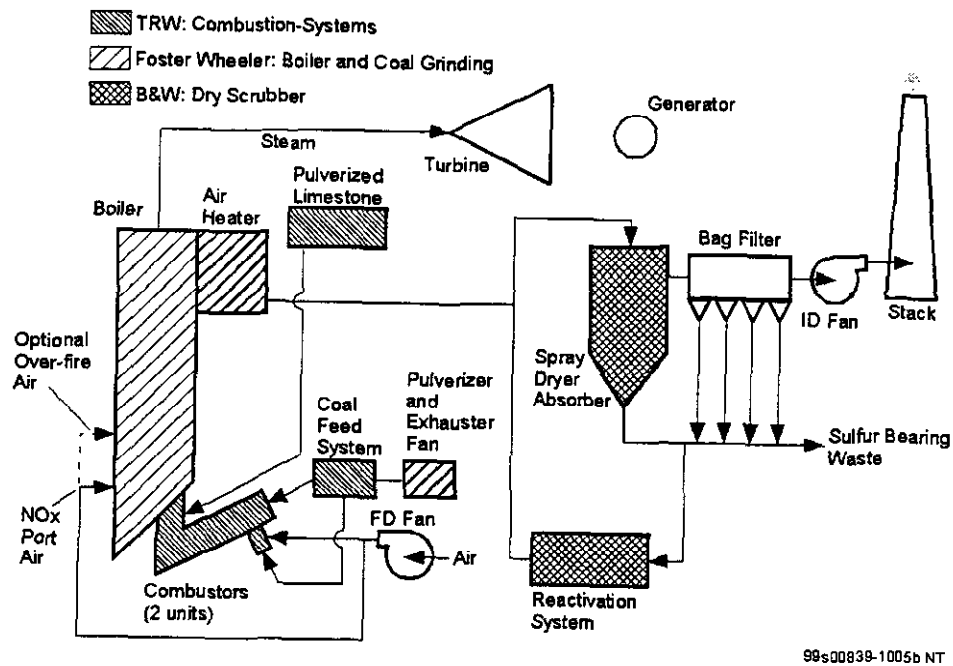


FIGURE 1 - HCCP INTEGRATED SYSTEM

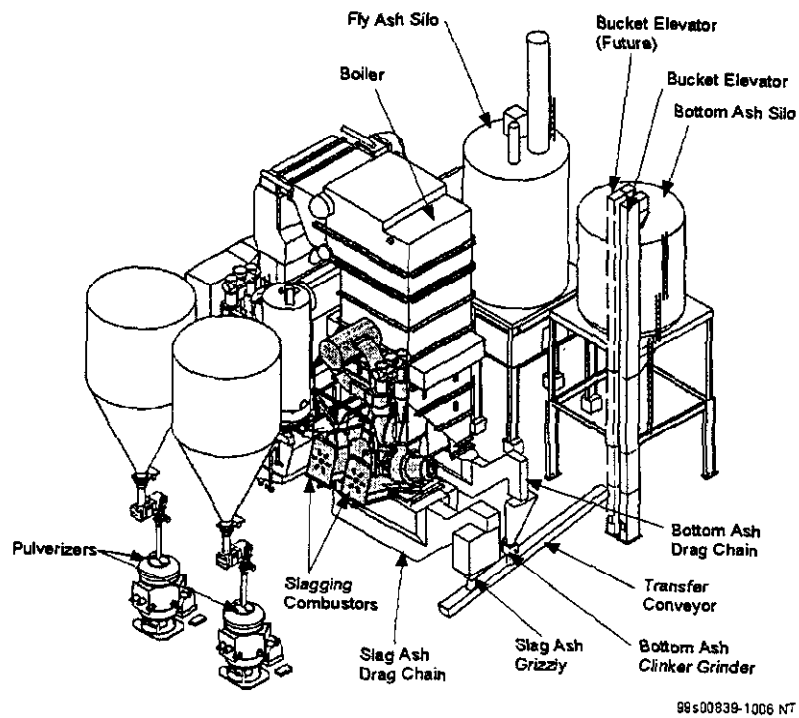


FIGURE 2 - ISOMETRIC VIEW OF HEALY BOILER ISLAND

Combustion System Description

Figure 3 illustrates an isometric view of one of the two 350 MMBtu/hr TRW (88 million kcal/hr) multistage slagging combustors designed for the HCCP. It consists of a precombustor, a slagging stage and a slag recovery section. The main chamber of the slagging stage is approximately 9 feet in diameter by 16 feet in length. The walls of the combustor were fabricated using tube-membrane construction, primarily with 1.5 inch SA213 T2 tubing and SA387 Grade 11 fin material. The combustors are cooled by a two-phase forced circulation system directly integrated with the boiler drum (1400 psia, 585 deg F). The twin combustors were fabricated at Foster Wheeler's facility in Dansville, NY, per TRW specification drawings and were transported to the plant in several subassemblies. The combustors are suspended from the boiler (top-supported).

A functional schematic of the combustion system is shown in Figure 4. Pulverized coal is injected in both the precombustor and slagging stage. The precombustor is used to boost the combustion air temperature from the air heater (from typically 500-700 deg F to 2300-3300 deg F) by burning 30 to 45% of the total pulverized coal flow rate. The precombustor is a vital component of the system because it controls the temperature and velocity of the hot combustion gases entering the slagging stage for optimum combustion and slag removal. It is designed to ensure stable, efficient combustion of a wide variety of coals, and to prevent slag freezing within the slagging stage while burning high ash fusion temperature coals under fuel rich conditions. Low volatility coals can be accommodated by firing a larger portion of the coal in the precombustor.

The high temperature, oxygen-rich combustion gases from the precombustor enter the slagging stage tangentially, generating a high velocity, high temperature confined vortex flow. The balance of the pulverized coal (55 to 70% of total) is injected through a multi-port injector at the head end of the slagging stage. The high gas temperature produced by the precombustor promotes a hot slagged surface on the interior of the slagging stage, which combined with the strong recirculation patterns, ensures stable ignition and combustion. The multi-port injector helps distribute the coal evenly for better coal/air mixing and combustion. The slagging stage is operated at fuel rich conditions at stoichiometric ratios typically in the range 0.7 to 0.9. Carbon conversion to combustion gases is maximized and NO_x emissions are minimized by controlling the temperature, gases and solids mixing and stoichiometric conditions in the slagging stage.

The precombustor, slagging stage and the slag recovery section are operated in a slagging mode, i.e., the coal ash melts to form a molten slag layer which coats the inside surfaces. The coal particles are combusted at a high enough temperature to melt the residual coal ash contained within each particle. Slag droplets are produced, which are centrifuged to the walls of the combustor, forming a self-replenishing slag layer. This slag layer is molten on the gas-side surface and frozen at the tubewall interface. The frozen slag layer protects the water-cooled metal body of the combustor from erosion, abrasion and corrosion, and also reduces the heat transferred to the water in the combustor body. The molten slag is transported along the walls by shear and gravity forces. The molten slag flows through a key slot, along the bottom to the slag tap opening located in the slag recovery section. Up to 90% of the slag is discharged through

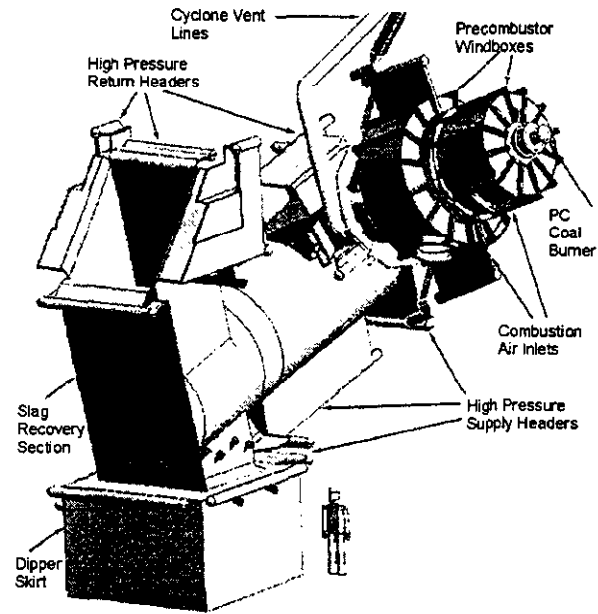


FIGURE 3 – ISOMETRIC VIEW OF ONE OF THE TWO 350 MMBTU/HR TRW SLAGGING COMBUSTORS

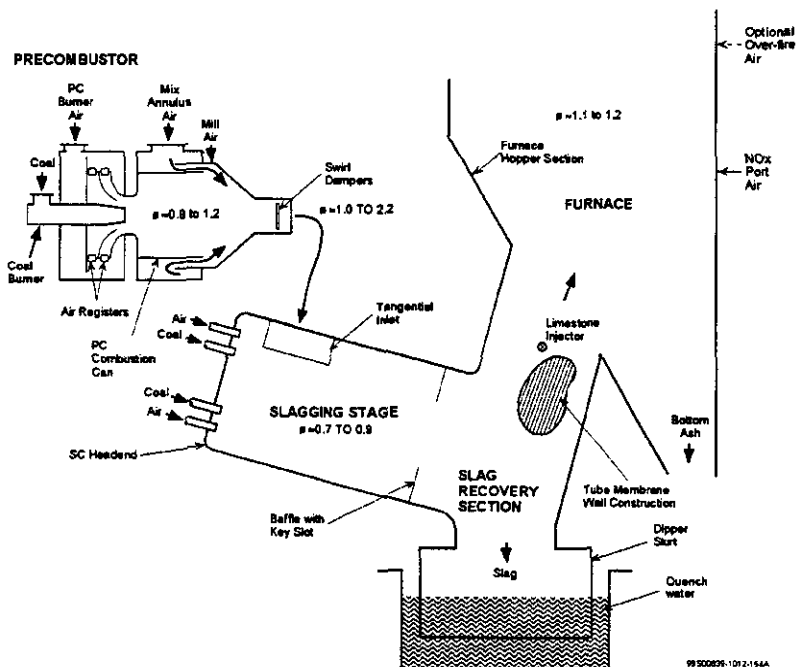


FIGURE 4 – FUNCTIONAL SCHEMATIC OF TRW COMBUSTION SYSTEM

the slag tap by gravity. A dipper skirt arrangement is used to provide a water seal for the system. The molten slag drops into the water, where it shatters upon contact and is rapidly quenched, yielding a granular glass-like product. The slag is removed from the slag tank by a drag chain conveyor.

Only 10 to 25% of the original coal ash enters the boiler. Because of the aerodynamics of the cyclonic slagging stage, the majority of this entrained slag will be molten droplets of less than 10 microns in size. As the fine slag droplets solidify at lower temperatures in the furnace, spherical shaped particles are formed that are expected to have lower fouling and erosion characteristics than conventional flyash particles, potentially increasing the life of the furnace and its convective tubes.

Emissions Control

Figure 5a presents typical Healy Clean Coal Project combustion side gases and solids flows when each combustor unit is operated at a firing rate of 315 MMBtu/hr on a Usibelli ROM and waste coal blend. The flow rates presented in Figure 5a are for the total plant, which includes two coal feed systems, one pulverized limestone feed system and two coal combustor systems. The coal feed system provides coal and a portion of the mill air directly to the precombustor and slagging combustor stages with the remaining mill air sent to the boiler NO_x ports. The warm combustion air from the air heater section is delivered to the boiler NO_x and over-fire air (OFA) ports and to both coal combustion stages. Low NO_x emissions are achieved when the slagging combustor gases are fuel rich / high temperature and the furnace combustion gases are fuel lean / low temperature.

In the slagging stage, a high flame temperature is achieved by preheating the combustion air in the air heater to as high a value as possible, and the stoichiometry is reduced to as low a level as possible without compromising on slagging and carbon conversion. As the combustion gases enter the furnace, the stoichiometry is still less than unity. The remaining air is added in the furnace either at the NO_x ports (as is done at HCCP) or at the OFA ports to complete the combustion at an overall stoichiometry of 1.1 to 1.2 (10 to 20% excess air). In the furnace, the addition of final air is delayed until the gas temperature is reduced by radiative cooling to the walls; this reduces the peak temperatures in the furnace. Also, excess air in the range of 10 to 20% is maintained not only to reduce NO_x but also to complete the combustion of any unburned carbon in the gases.

For mitigation of SO_2 emissions, the combustor offers the advantage of *in-situ* calcination of pulverized limestone (CaCO_3), which is injected in the upper region of the slag recovery section. The limestone particles are calcined in the furnace to highly reactive flash-calcined lime (CaO) particles. By the time these lime particles mix and move with the combustion products to the exit of the boiler, a portion of the SO_2 is absorbed to form calcium sulfate (CaSO_4). In the HCCP, the utilization of these flash-calcined lime particles is further enhanced by the back end flue gas desulfurization system specifically supplied by B&W/Joy/Niro. For the HCCP, when firing low

sulfur coal, approximately 5 to 20% sulfur capture takes place in the furnace when injecting pulverized limestone at a calcium-to-sulfur (Ca/S) molar ratio < 2 .

The flue gas desulfurization system, shown schematically in Figure 5b, is comprised primarily of a spray dryer absorber and pulse-jet baghouse. Auxiliary systems include a reagent (FCM) storage, preparation and feed system and an ash conveying system. Use of the FCM as the sole SO_2 scrubbing reagent is a unique feature of the process, resulting in significant cost savings over the conventional use of pebble lime as the reagent, which is typical for most dry flue gas desulfurization systems.

Combustion gases discharged from the air heater outlet of the unit is directed to a dedicated 100% capacity spray dryer absorber (SDA) and PulseFlo® pulse-jet baghouse system wherein SO_2 removal and particulate collection takes place.

The combustion gases enter the SDA module via the roof gas disperser, which distributes the incoming flue gas symmetrically around the rotary atomizer. The roof gas disperser promotes mixing (i.e. gas liquid contact) of the combustion gases and reagent slurry to promote drying, maximize SO_2 removal and minimize solids deposition inside the SDA. The SDA utilizes a NIRO F-350 rotary atomizer to atomize the feed slurry (i.e. a mixture of FCM, reaction products, flyash and water) into a fine spray and inject it into the incoming combustion gases.

The finely atomized feed slurry mixes with the combustion gases, resulting in the evaporation of water and the removal of SO_2 via chemical reaction with the hydrated lime component of the slurry. The chemical reactions that occur as the hydrated lime ($\text{Ca}(\text{OH})_2$) component of the FCM feed slurry reacts with the SO_2 produces reaction products in the form of calcium sulfite ($\text{CaSO}_3 \cdot \frac{1}{2} \text{H}_2\text{O}$) and calcium sulfate ($\text{CaSO}_4 \cdot 2\text{H}_2\text{O}$).

As the flue gas and feed slurry mixture pass through the spray dryer absorber, the concentration of the SO_2 is reduced substantially and the spray drying of the reagent slurry and reaction products is completed.

The combustion gases and entrained particles of calcium sulfite, calcium sulfate, unreacted reagent and flyash exit the SDA module into the PulseFlo® pulse-jet baghouse wherein the final step of the SO_2 and particulate removal processes takes place. The PulseFlo® pulse-jet baghouse removes $>99.9\%$ of the boilers exhaust solids, reaction products and recycled FCM before discharging the combustion gases to the stack.

Depending on percent ash removal in the combustor, coal sulfur content and Ca/S ratio, approximately 60-90% of the solids (i.e. reaction products, unreacted reagent, inerts, and flyash) collected in both the SDA module hopper and the pulse-jet baghouse hoppers is conveyed by the ash transport system to the flue gas cleaning system's FCM recycle surge bin. The remaining solids are rejected as waste.

Overall SO_2 removal efficiencies greater than 90% have been demonstrated when operating at furnace calcium to sulfur ratios in the range of 1.1 to 1.8.

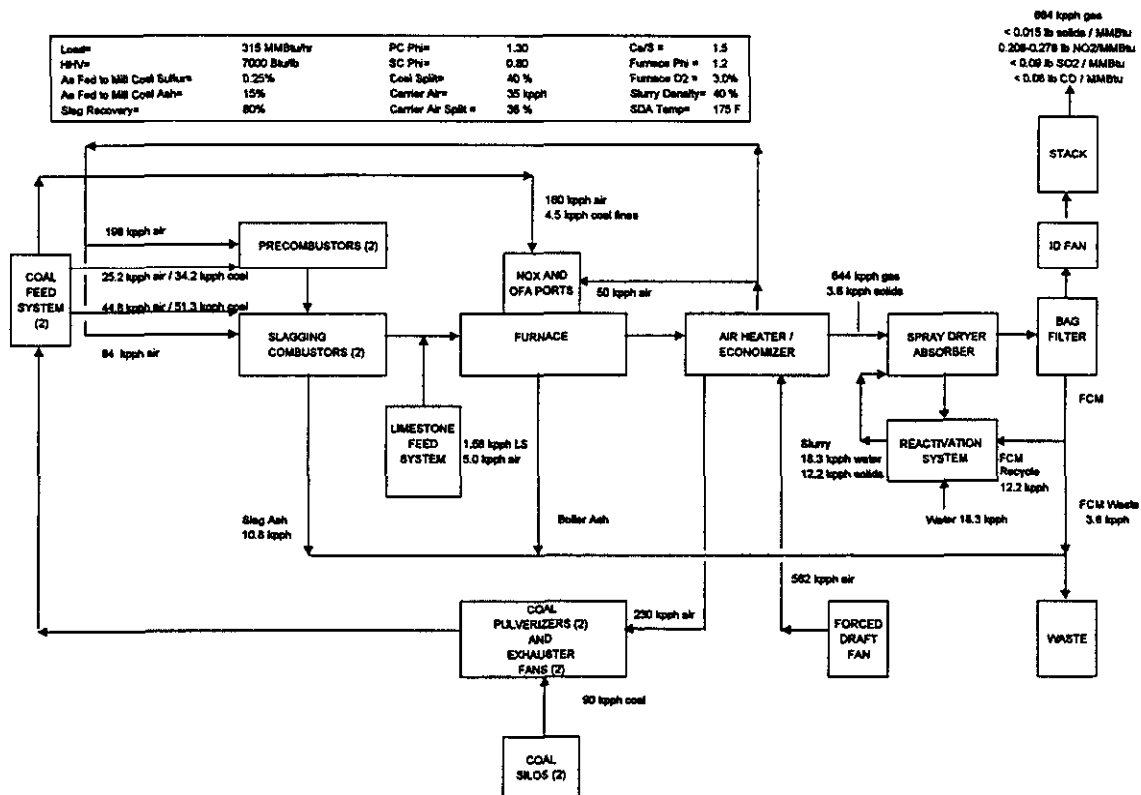


FIGURE 5A – HEALY CLEAN COAL PROJECT COMBUSTION SIDE FLOWS

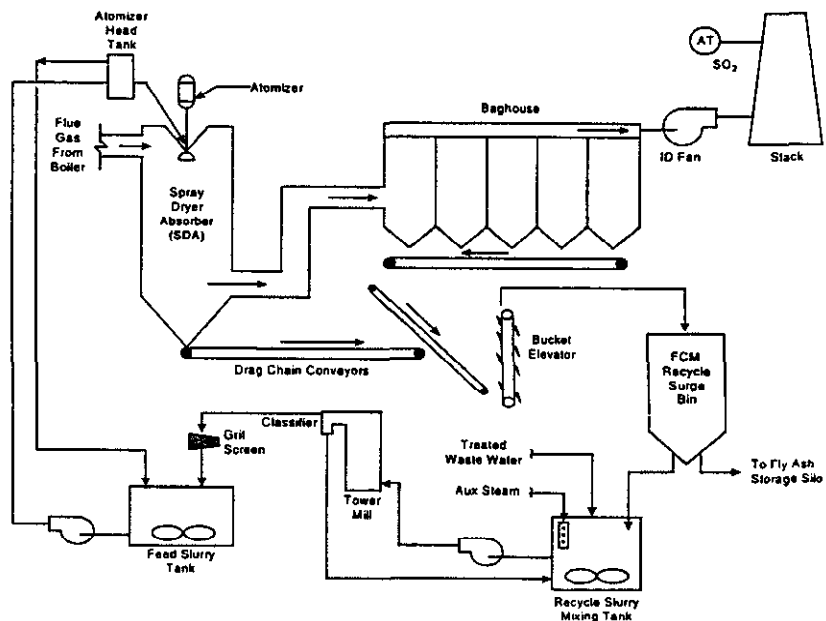


FIGURE 5B – FLUE GAS DESULFURIZATION SYSTEM

III. OPERATION AND PERFORMANCE RESULTS

Operation Summary

The 1998 Healy Demonstration Test Program consisted of several test activities, including Coal Firing Start-up Activities, Compliance Testing, Combustion System Characterization Testing, SDA Technology Characterization Testing, and Coal Blend Testing. The first 4 months of the Demonstration Test Program were dedicated to coal-firing start-up operations and focused on slowly bringing all plant systems on line while burning ROM coal at part-load operation. The plant reached full load for the first time in March 1998. Combustion System Characterization Testing was initiated in May 1998, concurrent with the initial firing of waste coal blends. The focus of the Combustion System Characterization Testing was to map the combustor performance characteristics over a broad range of operating conditions and hardware configurations. Shutdowns were incorporated into the test planning activities in order to inspect the combustor internal slagging characteristics as a function of the various hardware configurations and test conditions evaluated.

Overall in 1998, a total of 50 operational runs were conducted, accumulating 4,471 hours of cumulative coal burn time (not including oil-fired only start-up and shutdown time) on the Healy Coal Combustors. Of this total, 1,938 hours were on Run of Mine (ROM) coal and 2,533 hours were on ROM and waste coal blends. ROM coal was used primarily during: 1) plant coal firing start-up tests in the January to April, 1998 time frame, 2) emissions source testing in June 1998, and 3) when waste coal was not available from UCM in August and September 1998. The longest continuous run with ROM coal, conducted with both combustors at full load, was 431 hours (18 days). The longest continuous run with a blended coal, conducted primarily at part load with only one combustor in operation at full load, was 581 hours (24 days). During this test, the second combustor was shutdown following approximately 24 hours of coal-fired operation due to a vibration alarm on the mill Exhauster Fan as a result of non-uniform erosion of the fan blades.

In general, the composition of the ROM coal was fairly consistent from test-to-test, however, the blended coal composition varied significantly depending on the coal mining technique, the coal seam, and the type of coal blending technique used. Daily coal samples were taken by Golden Valley during loading operations, and were subsequently analyzed by Usibelli Coal Mine.

The overall range of coal properties tested from May through December, 1998, compared to the range of coal properties listed in the Design Specification, is as follows:

	<u>Design Basis</u>			<u>1998 Actuals (avg)</u> (May – Dec 1998)
	<u>Run of Mine</u>	<u>Performance</u>	<u>55/45 Blend</u>	
Higher Heating Value, (Btu/lb)	7815	6969	6874	6196 to 8271 (7507)
Vol. Matter, (%)	34.6	30.8	30.4	25.0 to 37.5 (35.1)
Fixed Carbon, (%)	30.9	27.5	27.2	24.1 to 30.9 (27.9)
Moisture, (%)	26.4	25.1	25.0	22.5 to 29.4 (25.9)
Ash, (%)	8.20	16.6	17.4	5.7 to 24.0 (11.1)
Sulfur, (%)	0.17	0.15	0.15	0.11 to 0.36 (0.18)
T ₂₅₀ (Deg F)	2228	2750	2800	2270 to 2900

As shown in the above table, the actual ranges in coal properties tested in 1998 were broader than the range indicated by the three different coal types listed in the Design Specification: Run-of-Mine, and two waste coal blends: 50% Waste / 50% ROM (also called "Performance Coal") and 55 % Waste / 45 % ROM.

Performance Results

This paper presents the results of coal-fired test operations from June 12 through December 21, 1998. During this period of time, approximately 3300 hours of plant thermal operation was accumulated, with approximately 3200 hours of coal-fired operation. The majority of test operations were at full load, 50 MW_e. The emission data presented includes all coal-fired operations during this period of time, including, in most cases, start-up and shutdown operations. Not included herein is emission data during: 1) January through June 11, 1998, which primarily consisted of plant start-up and shake down activities and was prior to certification of the Continuous Emissions Monitoring System (CEMS) and, 2) during oil-fired only operation.

Table 1 presents a summary of the Coal Combustion System and SDA performance goals, New Source Performance Standards (NSPS), and HCCP Air Quality Permit requirements compared to the performance results demonstrated during coal-fired test operations from June 12 through December 21, 1998. As noted above, the emission data presented in the table includes all coal-fired operations during June 12 through December 21, 1998, including in most cases, coal-fired start-up and shutdown operations, but does not include oil-fired only operations. The average values for NO_x, SO₂, CO, and Opacity listed in Table 1 were determined by averaging the emission data recorded on the plant data recording system (referred to as ODMS) during the approximately 3200 hours of coal-fired operation from June 12, 1998 through December 21, 1998. The NO_x emission data presented in the table is based on a 30-day rolling average, whereas the SO₂, CO, and opacity data averages are based on 30-minute averages. As shown, the performance results for NO_x, SO₂, and CO demonstrated during coal-fired operations from June 12 through December 21, 1998, met or exceeded all performance goals. As noted in the table, during 1998, the opacity and particulate matter were higher than anticipated due to a problem

with the baghouse. Following modification of the baghouse in December 1998, the opacity and particulate matter emissions are meeting performance goals.

TABLE 1 – HCCP PERFORMANCE GOALS AND RESULTS

PARAMETER	New Source Performance Standards (NSPS) [1]	HCCP AIR QUALITY PERMIT	CONTRACT GOALS	DEMONSTRATED IN 1998 (June - December, 1998)	
				RANGE	TYPICAL
NOX	0.5 lb/MMBtu (prior to 7/97) 0.15 lb/MMBtu (modified after 7/97) 1.6 lb/MMBtu (new plant after 7/97)	0.350 lb/MMBtu (30 day rolling average)	< 0.35 lb/MMBtu	0.208-0.278 lb/MMBtu 30-day rolling ave. [9], [10]	0.245 lb/MMBtu 30-day rolling ave. [9], [11]
CO	Dependent on ambient CO levels in local region (Title V of 1990 CAAA)	0.20 lb/MMBtu, (hourly average) (202 ppm CO @ 3.0% O ₂)	< 200 ppm (dry basis) at 3.5% O ₂ (dry basis) [2] <208 ppm CO @ 3.0% O ₂	<130 ppm at 3.0% O ₂ [5], [8]	30-40 ppm at 3.0% O ₂ 0.036 lb/MMBtu [5], [8]
SO ₂	80 % removal and less than 1.2 lb/MMBtu 70% removal when emissions are less than 0.60 lb/MMBtu	0.085 lb/MMBtu, (annual average) 0.10 lb/MMBtu, (3-hour average) 65.8 lb/hr max, (3-hour average)	70 % Removal (minimum) 79.6 lb/hr SO ₂ (maximum)	< 0.09 lb/MMBtu (<35 ppm @ 3% O ₂) [5], [8]	0.038 lb/MMBtu (15 ppm @ 3% O ₂) (25 lb/hr) [5], [8]
OPACITY	20% Opacity (5 min. average)	20% Opacity, (3 min average) 27% Opacity (one 6 min period per hour)	20% Opacity, 3 min average	<10 % Opacity [6]	5.6% Opacity (Jun - Dec 1998) [5],[15] 2.3% Opacity (1999) [15]
PARTICULATE MATTER	0.03 lb/MMBtu	0.02 lb/MMBtu, (hourly average)	0.015 lb/MMBtu		0.0047 lb/MMBtu (1999) [14], [15]
CARBON BURNOUT	NA	NA	> 96% at 100% MCR for Part., ROM, and 55/45 Blend [5] >98% at 100% MCR for Waste Coal	NA	99.7% [4]
SLAG RECOVERY	NA	NA	> 70% at 100% MCR for all coals [5]	78-87% [7]	83% [7]
NET POWER PRODUCTION	NA	NA	50 MW _e for all coals	NA	50-65 MW _e [12],[13]

NOTES
 [1] From 40CFR60.40a - 40CFR60.48b; New NO_x Standards based on 62 FR 36948
 [2] From minimum to 100% MCR (Maximum Continuous Firing Rate)
 [3] 100% MCR for Performance Coal is 315 MMBtu/hr, ROM Coal is 306 MMBtu/hr, Waste Coal is 322 MMBtu/hr, 55/45 Waste/ROM Coal is 316 MMBtu/hr
 [4] Measured for one test based upon slag and flyash carbon contents
 [5] Average of available 30 min. (average) test data, June 12, 1998 to December 21, 1998 (total of 3100 hours of run time)
 [6] 95% of CO, SO₂, and opacity data are observed to be less than these reported value (using available 30 min average test data)
 [7] Slag weight corrected for 6% moisture content.
 [8] Data corrected to 3% O₂
 [9] 30-day rolling average determined from available 30 min (average) test data, June 12, 1998 to December 21, 1998, total of 3100 hours (5480 data points).
 30-day rolling average only includes days in which power was generated.
 [10] Represents minimum and maximum of 30-day rolling average data described in Note [9]
 [11] Represents the average of 30-day rolling average data described in Note [9]
 [12] Nominal power set point from April through September, 1998 was 60-62 MW_e (gross), 53-55 MW_e (net);
 [13] Nominal power set point in November and December, 1998 was 57 MW_e (gross), 50 MW_e (net)
 [14] Based on independent particulate matter testing performed on March 10-11, 1999 by Hest, Morgan & Hudson
 [15] Opacity and particulate matter emissions during 1998 were higher than expected due to a problem with premature baghouse filter bag failure, which was corrected in 1999

Figure 6 plots the key plant parameters (power, boiler %O₂) and stack emissions (NO_x, SO₂) as a function of time for 13 days of an 18-day continuous run conducted with both combustors at full load burning ROM coal from June 8 through June 26, 1998 (remaining data from test run not available). Figure 7 shows the same parameters for a 24-day continuous test run from September 27 to October 21, 1998 on a waste coal blend conducted primarily at part load with only one combustor in operation. The key statistics from these extended test runs are provided below:

	<u>Run of Mine</u>	<u>Waste Coal Blends</u>
Test Period	6/12/98 – 6/25/98	9/27/98 – 10/21/98
Test Hours	312 hours	580 hours
Average NO _x in Exhaust	0.233 lb/MM Btu	0.204 lb/MMBtu
Average SO ₂ in Exhaust	0.030 lb/MM Btu	0.035 lb/MMBtu
Average O ₂ in Boiler	3.50 %	6.75 %
Average Gross MW _e	59.9 MW _e (2 combustors at full load)	29.8 MW _e (1 Combustor at full load)

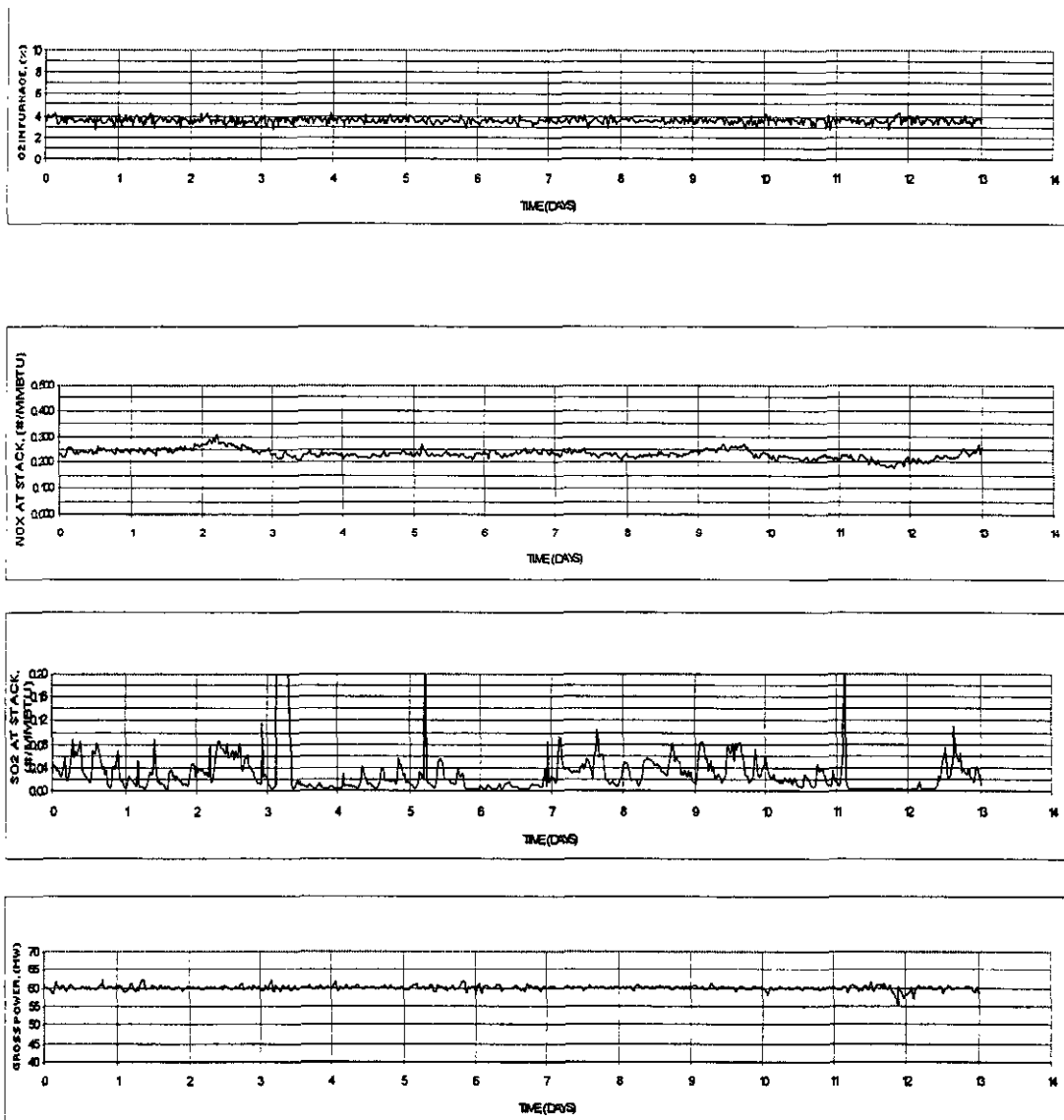


FIGURE 6 - HCCP EMISSIONS DURING 13 DAYS OF CONTINUOUS OPERATION WITH RUN OF MINE COAL (BOILER AT FULL LOAD - 2 COMBUSTORS), JUNE 12, 1998 TO JUNE 25, 1998

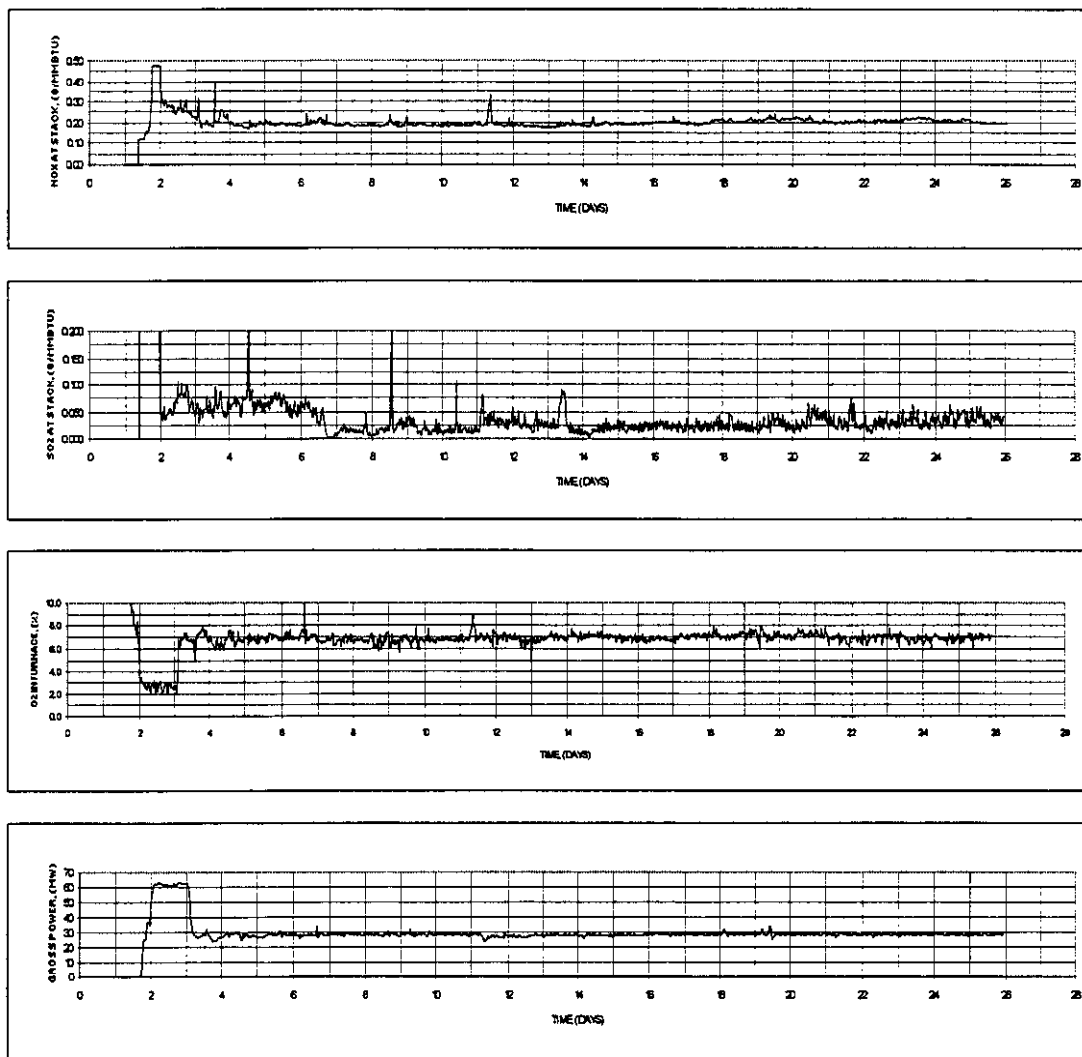


FIGURE 7 – HCCP EMISSIONS DURING 24 DAYS OF CONTINUOUS OPERATION WITH WASTE COAL BLEND (BOILER AT PART LOAD – 1 COMBUSTOR), SEPT. 27, 1998 TO OCT. 21, 1998

During the continuous runs, the plant produced 58-62 MW_e with two combustors in service and 28-30 MW_e with only one combustor in service. As illustrated by the stack emission trends indicated in Figure 6, the emission levels of NO_x and SO₂ were very consistent during the steady-state portion of the test. No problems were experienced with the Run of Mine coals or with any of the coal blends in the slagging combustor stage. In early testing with waste coal blends with heating values below 7400 Btu/lb in combination with wide coal property variations (particularly heating value, ash content, ash T₂₅₀), slag freezing in specific areas of one or both of the two operating precombustors would occur over a period of several days. Several secondary air injection modifications were evaluated in order to minimize this slag freezing phenomena: 1) Improve secondary air mixing by injecting the air into the core flow of the precombustor combustion products through high velocity discrete air jets, and 2) Relocating a portion of the Secondary Air from the precombustor to the headend of the slagging stage. Ultimately, precombustor slag freezing was minimized by relocating the secondary air injection to the slagging stage and by transferring the excess mill air (i.e., the additional mill air not required for coal transport) to the boiler after start up. These changes not only eliminated the mixing of air downstream of the precombustor combustion chamber, but it also effectively increased the precombustor operating temperature to the 3200-3500 deg F level. During 1999, additional adjustments to the precombustor coal burner configuration (e.g., adjustment of coal fines injection velocity and inner and outer air register settings) were made in order to broaden the operating envelope when burning ROM/Waste blend coal.

As the Demonstration Test Program continues, additional emission data will be collected to characterize the integrated HCCP performance over longer operational periods. Longer-term operation planned for 1999 will provide the opportunity to evaluate and optimize the HCCP emission performance over a wider range of ROM/Waste blend coals and limestone characteristics as a function of boiler load and boiler excess air. This will enable a more accurate projection of the expected HCCP performance during future long-term commercial operations.

Analytical Model Comparisons

During May 1998 (prior to certification of the continuous emission monitoring equipment), parametric tests were performed in order to: 1) determine the boundaries for key operating variables (e.g., slagging combustor stoichiometry) and 2) provide a basis for comparison to analytical model predictions of the HCCP combustor performance. Although this data was not presented in the previous section, which discusses emission performance results, it is presented in this section in order to show a comparison between “predicted” performance and “actual” performance.

Figure 8 presents the TRW NO_x model predictions for the Healy combustor as a function of slagging combustor stage stoichiometry. Superimposed on the model are data points from tests conducted at Healy with a ROM/Waste blend coal during May 1998 when the slagging stage stoichiometry was varied in order to map NO_x as a function of stoichiometry. In general, good agreement was obtained between model predictions and actual test results. The combustor stoichiometry was observed to be the most important combustor operating parameter for NO_x

control, while changes to the combustor coal split and air split had secondary effects. Most of the full load tests were conducted at combustor air / fuel stoichiometries between 0.80 and 0.85. This stoichiometric range was selected since it yielded low NO_x emissions while still maintaining high slag recovery, high carbon burnout and low CO emissions. According to the model, it may be possible to further reduce NO_x through additional optimization of the combustor operating conditions. NO_x data obtained during 1998 also indicated that the furnace O_2 may have been higher than optimal (typical range was 3.5% to 4.5%) for minimum NO_x formation.

Figure 9 presents the TRW model for in-situ sulfur capture in the furnace as a function of Ca/S ratio and coal sulfur content. Superimposed on the model are data points from tests conducted at Healy with a waste blend coal with 0.3% sulfur coal and 74 micron median size limestone injection, during May 1998. During this period of time, the limestone feeder was not accurately calibrated, and, therefore, the limestone flowrate was determined based on several grab samples taken during the test. These grab samples were used to develop a correlation between limestone feeder belt speed and limestone flowrate. The sulfur reduction shown for the Healy test data was determined by comparing the baseline SO_2 emissions at the furnace exit without any limestone flowrate to the SO_2 emissions at the furnace exit with limestone flowrate at various Ca/S ratios. Also included in Figure 9, for reference, is the data from the industrial size combustor tests in Cleveland and CTS where fine sized (7-25 micron) limestone was used with high sulfur coals (~3%). Due to use of low sulfur coal at Healy, the combustors and furnace are primarily being used for calcination of the limestone and only a relatively low level of sulfur capture occurs in the furnace. At HCCP, the utilization of the in-situ flash-calcined lime particles is further enhanced by the back end flue gas desulfurization system and baghouse, which results in up to 99% sulfur capture. However, based on the data from the industrial size combustor tests, during operation with higher sulfur coals, additional sulfur capture within the furnace is expected for a given Ca/S ratio.

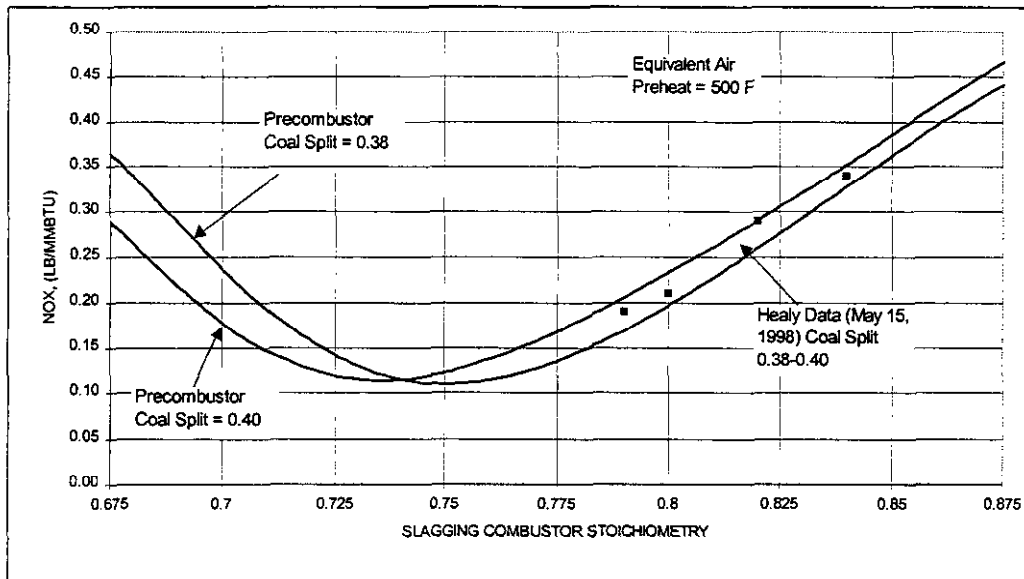


FIGURE 8 – COMPARISON OF HEALY NO_x DATA WITH TRW NO_x MODEL (300 MMBTU/HR)

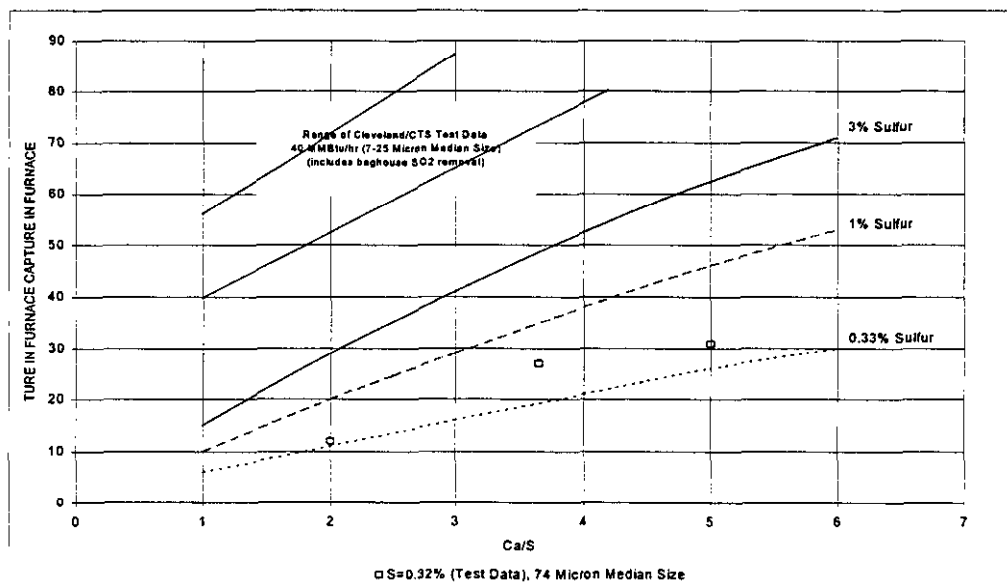


FIGURE 9 – COMPARISON OF HEALY FURNACE SULFUR CAPTURE WITH TRW SULFUR CAPTURE MODEL PREDICTIONS

Figure 10 presents preliminary slag recovery data. A “total” slag recovery value was determined for a cumulative test period covering 45 days, during which 4 tests were conducted (including 4 start-up and shutdown periods). Slag recovery was determined to be approximately 80-85% over this 45-day period, based on ash hopper load cell measurements. This value includes bottom ash,

which is estimated to contribute less than 5% to the total ash capture. This reduction in the quantity of coal ash entering the furnace has several benefits, including: 1) reduction in ash loading through the boiler convective pass, 2) reduction in ash loading on the baghouse bags, and 3) reduction in total ash loading to the SDA which reduces the limestone flow requirements.

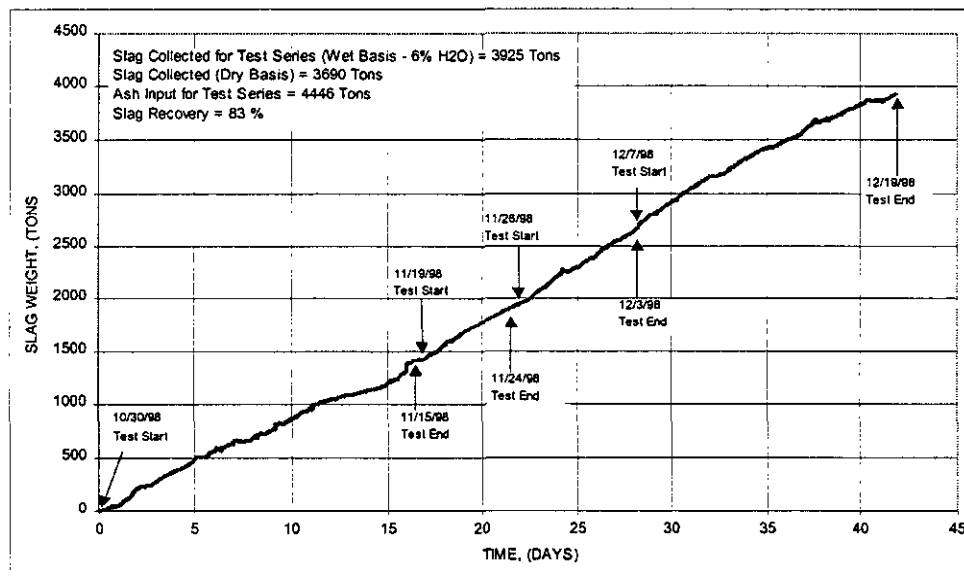


FIGURE 10 – DEMONSTRATION OF SLAG COLLECTION OVER 4 TESTS OVER A 42 DAY PERIOD BASED ON SLAG ASH HOPPER LOAD CELL READINGS

IV. FUTURE

HCCP Improvements

During the 1998 HCCP Demonstration Test Program, some site-specific integrated plant operational and/or hardware durability problem areas were identified and are currently being addressed. The following table summarizes the site-specific issues identified and the planned resolutions being implemented.

Problem Area	Planned Resolution
Ash/slag accumulation on the Furnace Hopper Slope; Ash/slag fall into the water-filled slag hopper results in trips on high Furnace pressure	Installation of a water lance on the Furnace Hopper Slope to mitigate ash accumulation in this region
Slag on internal surfaces occasionally obscuring flame scanner view angle	Integrate slag rodding capability on all flame scanner ports; provide additional scanner locations to ensure continuous flame monitoring
Erosion of blades and outer casing on mill exhausters fans	Incorporate improved erosion resistant materials on blades and outer casing; establish inspection program and provide spare materials

Future Tests

In order to further validate combustor scaling methodology and TRW-developed computer models for NO_x emissions and in-furnace sulfur capture, it would be useful to test the Utility-scale TRW Coal Combustion System with (1) high sulfur coal (2-3%) to fully validate the TRW in-furnace SO_2 capture model shown in Figure 9 and (2) ammonia (or urea) injection in the combustion stage to attempt to demonstrate the capability of this coal combustion system to meet the latest NSPS NO_x requirements (1.60 lb NO_x / MWh for new plants, 0.15 lb/MMBtu for modifications) as predicted by the TRW model presented in Figure 11. This analytical model of the NO_x reduction process was developed by TRW and is anchored to available experimental data. As shown, at the utility-scale, NO_x reductions down to the 0.10 to 0.15 lb/MMBtu level appear to be achievable at a NH_3 : NO molar ratio of 2 to 3.

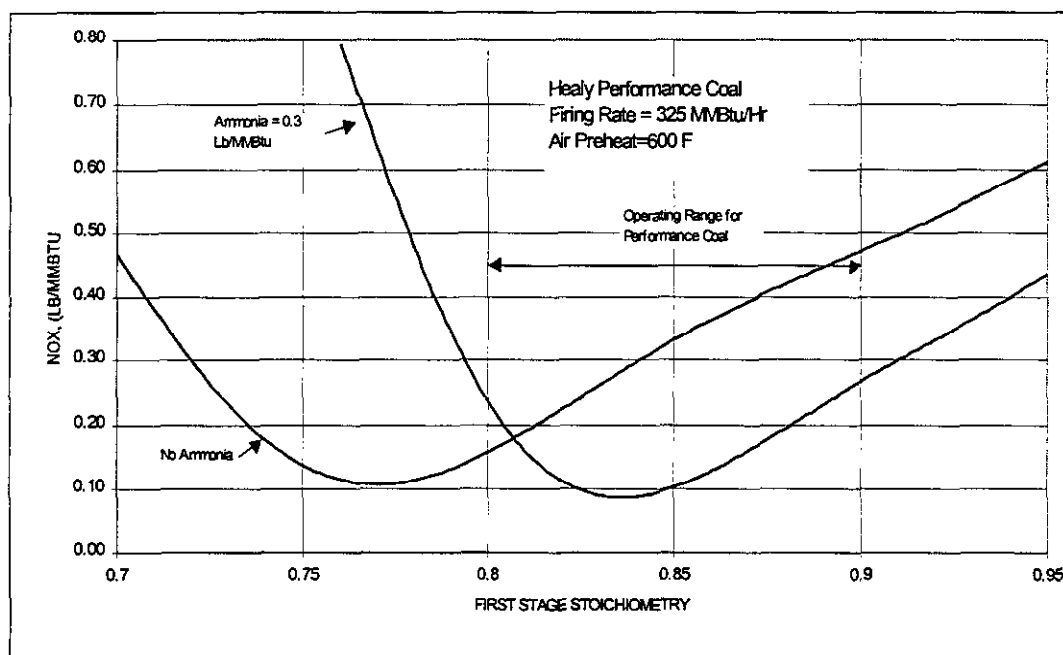


FIGURE 11 – PREDICTED NO_x LEVELS IN HEALY COMBUSTOR WITH AND WITHOUT AMMONIA INJECTION

Future Design Improvements

In addition to the operational demonstration and the demonstration of the ability to control emissions of NO_x , SO_2 , CO, and particulate matter, experience gained at the utility size will lead to significant design and cost reduction improvements. The Healy Clean Coal Combustion System was conservatively designed in many areas because of its first-of-a-kind status and site-specific requirements. Some of the system design changes planned for the next generation coal combustion systems are:

Current System Design	Future Design Improvements
High pressure coolant circulation pumps	Natural circulation coolant system
Coal mill exhausters fans	Direct coal feed from pressurized Mill
Multi-stage air injection in the precombustor (eliminated December 1998)	Single stage air injection in the precombustor

The above design changes and basic improvements in combustor system design will lead to lower system costs, increased reliability, and more economic operation. The coal combustor is designed using standard boiler design technology and procedures/processes. It was estimated by the boilermakers doing the actual Healy coal combustor fabrication that the costs would be halved on the next units produced. TRW is now in the process of updating these costs using the knowledge gained in manufacturing the Healy system, experience and data from the 1998 testing and better estimates from future hardware manufacturers and component suppliers.

Commercial Applications

The TRW Coal Combustion System offers the capability of using a wide variety of coals (including hard to burn coals) because of its multistage operating flexibility and combustion process control. A reduction in size and a life extension of boilers that incorporates the TRW Coal Combustor occurs because of the high carbon conversion and reduced solids loading in the combustion gases entering the boiler. The TRW Coal Combustor provides the capability of converting oil and gas fired boilers to coal and to eliminate oil or gas co-firing used with certain hard to burn coals. In parts of the world, the TRW Coal Combustor is attractive based on its ability to burn low grade local coals with low NO_x and particulates emissions while removing 50% of the SO₂ in the furnace and baghouse. The scrubbing of the remaining SO₂ could be added at a later time after the system is on line generating revenue. This system can be used to retrofit a wide variety of existing size boilers and the precombustor and slagging stage head end could be installed on existing cyclone furnaces to bring them into NO_x compliance and allow them to also burn local high sulfur coals.

V. ACKNOWLEDGEMENTS

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STATUS OF THE LAKELAND MCINTOSH UNIT 4 ADVANCED CIRCULATING FLUIDIZED BED COMBINED CYCLE DEMONSTRATION PROJECT

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ABSTRACT

In December 1997, the City of Lakeland, Florida, signed a Cooperative Agreement with the U.S. Department of Energy (DOE) that will facilitate the demonstration of the Advanced Circulating Fluidized-Bed Combined Cycle technology being developed by Foster Wheeler. The project will be conducted under the DOE Clean Coal Technology Program at the City of Lakeland's McIntosh Power Station in Lakeland, Florida. In August 1998, Lakeland authorized Foster Wheeler to begin the preliminary engineering and permitting support of the demonstration plant.

The Lakeland McIntosh Unit 4 Project is a nominal 240 MWe combined cycle plant that integrates the partial gasification of coal with pressurized circulating fluidized bed (PCFB) combustion. The partial gasification process produces a low Btu syngas and a coal char residue. The latter is burned in a PCFB boiler producing steam for a steam turbine and hot vitiated air/flue gas for a high efficiency gas turbine. The syngas, in turn, is burned in the gas turbine topping combustor (topping combustion) to heat the vitiated air to over 2300°F for expansion through the turbine.

The plant is designed to burn both low and high sulfur coal and incorporates a Siemens V64.3 gas turbine with a 2400 psig/1000°F/1000°F steam turbine. This paper describes the demonstration plant and identifies its design status.

I. INTRODUCTION

The City of Lakeland operates two power stations totaling approximately 820 MWe of generating capacity of which about 80% is wholly owned by Lakeland Electric Utilities. The McIntosh Station on the North side of Lake Parker is the larger of the two with approximately 590 MWe of generating capacity; the smaller Larsen Station on the South side of the lake has about 230 MWe of generating capacity.

The City of Lakeland has experienced and is forecasting steady load growth within its system of approximately 15 MWe per year; this will result in a capacity shortfall of approximately 60 MWe by the year 2000. In addition, Lakeland expects to retire 70 MWe of inefficient generating capacity. Faced with this load growth and anticipated retirement of older units, Lakeland plans to add approximately 200-250 MWe of new generating capacity.

To help meet their new power generation requirement, Lakeland plans to build a nominal 240 MWe plant utilizing Foster Wheeler's Advanced Circulating Fluidized Bed Combined Cycle (ACFBCC) technology. The plant integrates the partial gasification of coal with pressurized circulating fluidized bed (PCFB) combustion. The partial gasification process produces a low Btu syngas and a coal char residue. The latter is burned in a PCFB boiler to produce steam for a steam turbine and hot vitiated air/flue gas for a gas turbine. The syngas in turn is burned in the gas turbine (topping combustion) to heat the vitiated air to over 2300°F.

The McIntosh Unit 4 PCFB plant will be constructed on undeveloped land located adjacent to the existing McIntosh Unit 3. The plant will be designed to burn a range of coals including both the current Eastern Kentucky coal burned in the conventional pulverized coal fired Unit 3 as well as lower priced, high ash, high sulfur coals that are available on the open market. Limestone will be procured from Florida sources while the ash will be disposed in landfill or marketed.

The plant will be funded in part through the U.S. Department of Energy (DOE) Clean Coal Technology (CCT) Program. The DOE funding results from a combination of two previous Clean Coal awards: the DMEC-1 PCFB Repowering Project selected under Round III and the Four Rivers Energy Modernization Project (FREMP) selected under Round V. The DMEC-1 project was intended to demonstrate non-topping PCFB technology (gas turbine temperature is essentially the same as the PCFB temperature), while the FREMP project was planned to demonstrate Topped PCFB technology (gas turbine inlet temperature is markedly higher than the PCFB temperature).

II. PROCESS DESCRIPTION

A non-topped PCFB plant is a combined cycle power generation system employing gas and steam turbines and combusting solid fossil fuel in a PCFB boiler. Tubes contained in the PCFB generate, superheat, and reheat steam for use with the most advanced steam turbines (Rankine cycle) and the hot, pressurized combustion flue gas/vitiated air emanating from the PCFB in turn

can drive a gas turbine (Brayton cycle) for additional power generation. A non-topped PCFB plant can achieve thermal efficiencies in excess of 40 percent (HHV) and have a levelized busbar cost of electricity below any competing coal technology. In addition to the economic benefits, the built-in feature of environmental control (SO_2 and NO_x) in the combustion process eliminates the need for any external gas clean up such as scrubbers. A PCFB can also burn a much wider range of coals than a pulverized-coal-fired boiler. PCFB combined-cycle power plants offer real economic incentives for low cost electric power generation in an environmentally acceptable manner, while burning a wide range of low cost, abundant coals.

Figure 1 represents a simplified schematic of Foster Wheeler's Non-Topped PCFB Combined Cycle. Combustion and fluidizing air is supplied from the compressor section of the gas turbine to the PCFB combustor located inside a pressure vessel. Coal and sorbent (usually limestone) are mixed with water into a paste which is pumped into the combustion chamber using reciprocating pumps commonly used in the concrete industry. The same type of pumps have been successfully proven in a number of pressurized fluidized bed combustion plants and facilities around the world. The limestone sorbent captures sulfur in situ as sulfur dioxide, and nitrogen oxides are controlled by temperature and pressure.

Combustion takes place in the fluidized bed combustor at a temperature of approximately 1550 - 1600°F and, depending on the gas turbine used, typically 10 to 16 atmospheres. Particulate matter entrained in the flue gas exiting the combustor is removed using cyclones and ceramic barrier filters, such as a Siemens Westinghouse ceramic candle type Hot Gas Particulate Filter System (HGPFS), located between the PCFB and gas turbine. The high temperature, high pressure HGPFS is similar to that tested for 6000 hours at the American Electric Power PFBC Demonstration facility (Tidd) in Brilliant, Ohio ^[1]. Modules of this type of filter system have also undergone extensive testing at Foster Wheeler's PFB pilot plants in Livingston, New Jersey, and in Karhula, Finland ^{[2] [3]}, and in the Wilsonville Power Systems Development Facility operated by Southern Company Services for the DOE ^[4]. In addition to protecting the gas turbine from erosion, the HGPFS eliminates the need for any particulate removal at the stack thereby eliminating the need for a back-end electrostatic precipitator (ESP) or baghouse.

The hot gas cleaned by the filter system expands through the gas turbine, exhausts to a heat recovery unit, and vents to a stack. The heat recovered from both the combustor and the heat recovery unit is used to generate, superheat and reheat steam for use in the steam turbine. Approximately 15 to 25% of the total power produced is generated in the gas turbine, and the balance is generated in the steam turbine.

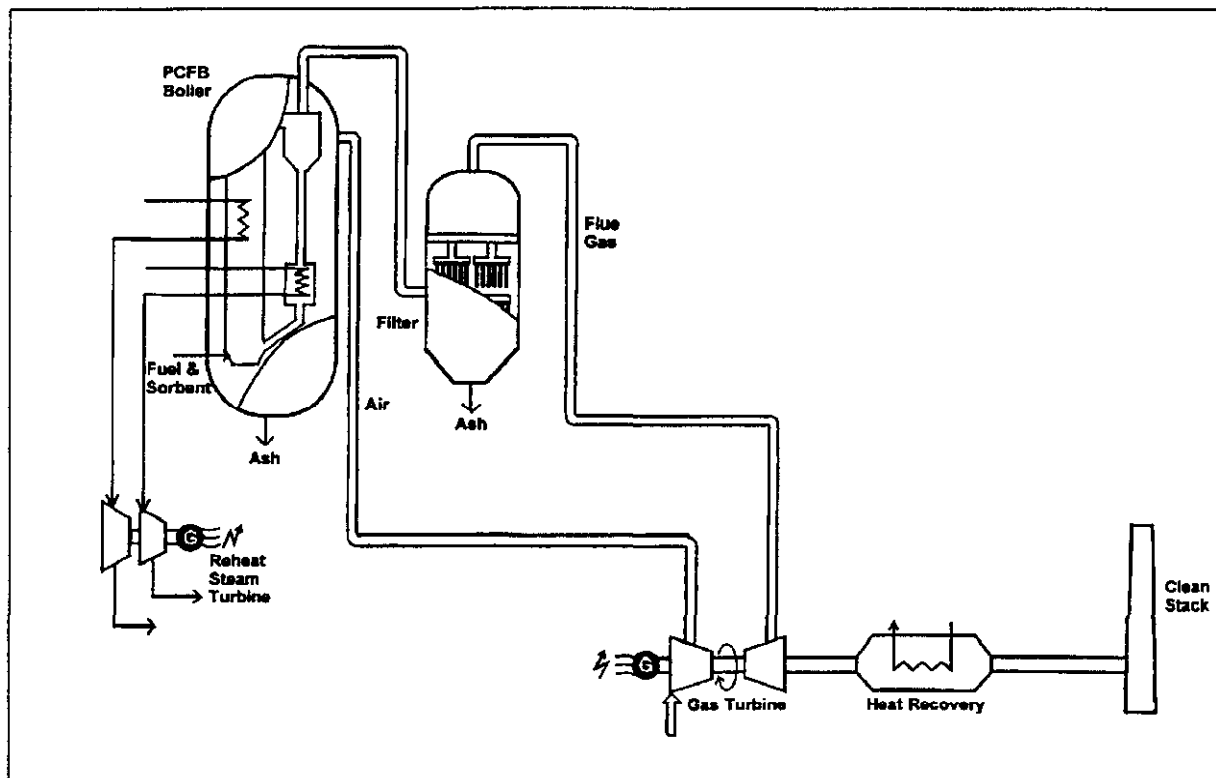


Figure 1 Foster Wheeler Non-Topped PCFB Cycle

Figure 2 shows a simplified schematic of Foster Wheeler's Topped PCFB/Advanced Circulating Fluidized Bed Combined Cycle Plant. ACFBCC technology integrates a carbonizer island and gas turbine topping combustor in the PCFB cycle. The additional components allow the firing temperature of the gas turbine to be increased to state-of-the-art levels; this is achieved by the combustion of a coal derived, low-Btu syngas produced in the carbonizer and fired in the gas turbine topping combustor. As a result, the gas turbine power output increases relative to the steam turbine, thereby increasing the plant efficiency to levels greater than 46 percent.

steam turbine power can be almost tripled. Although the plant efficiency decreases, the increased power is relatively inexpensive. In addition, the carbonizer char transferring to the PCFB and syngas and flue gas proceeding to the gas turbine topping combustor can be cooled with boiler feed water. Gas cooling extends candle filter life, minimizes ash bridging potential in the filters, reduces gas turbine hot corrosion risks (alkali vapors condense on the particulate being removed by the filters), and eases material selections for valves, piping, etc. The ACFBCC Plant can be designed to operate, depending on the utility's preference, at either extreme (peak efficiency vs. maximum power output) or any point in between.

The ACFBCC Plant with its large gas turbine output is envisioned for larger size plants (250 to 500 MWe) whereas the Non-Topped PCFB with its smaller gas turbine is ideal for smaller size plants (100 to 400 MWe) as well as repowerings.

III. LAKELAND PLANT DESCRIPTION

Foster Wheeler has previously presented ^[5] performance data for the demonstration plant that showed a 189 MWe net power output with 42.8 percent efficiency (coal higher heating value basis). This efficiency is less than the cycle's full potential because a medium rather than a large size gas turbine is being used. The plant design at that point in time incorporated a 401F gas turbine; this was an 87 MWe gas turbine being developed by Westinghouse. In August 1998, Siemens AG purchased the Westinghouse Power Generation Division of CBS, and it has been decided to replace the 401F with the 60 MWe Siemens V64.3 gas turbine. Although the latter is well proven (over 30 in operation), it is a smaller machine furnishing less air and less power. To offset the lower gas turbine power output (30 MWe) and provide Lakeland with the additional power it desires, the steam turbine is being increased in size to 200 MWe. The PCFB, however, cannot furnish the additional steam required by the enlarged steam turbine because there is not enough air to support additional direct coal feed; instead, we will burn carbonizer char and coal in the gas turbine heat recovery steam generator to generate the additional steam. With the steam turbine to gas turbine power ratio increased to approximately three to one, the plant has become a maximum power output rather than a peak efficiency plant. As of the writing of this paper (January 1999) performance data for the V64.3 modified to operate with Westinghouse Multi Annular Swirl Burners had not yet been finalized. Our preliminary calculations, however, indicate the plant will have a net output of 238 MWe with an efficiency greater than 40 percent (HHV basis), and it will demonstrate the key features of our ACFBCC technology.

IV. PROJECT SCHEDULE

In August 1998 the City of Lakeland authorized Foster Wheeler to proceed with the preliminary engineering and permitting support of the plant once certified gas turbine performance data becomes available. In the meantime, Lakeland is in the process of initiating permitting and licensing activities that are now expected to take two years.

The project schedule is shown in Figure 3. The design of the facility (Phase 1) will coincide with and continue until the permitting process is completed. Thereafter, Phase 2 will begin with the general release for fabrication and construction, and last for 30 months to mechanical completion. Phase 3 will begin with the start up of the first (non-topping PCFB) demonstration. After up to 12 months of operation and testing, the gas turbine topping combustor will begin operation with natural gas. A similar 12 month test period is envisioned after which the carbonizer leg of the plant will be activated and the second (topping PCFB) demonstration begun. Topping PCFB operation and testing will continue for two years, after which the plant will be released to Lakeland for commercial operation.

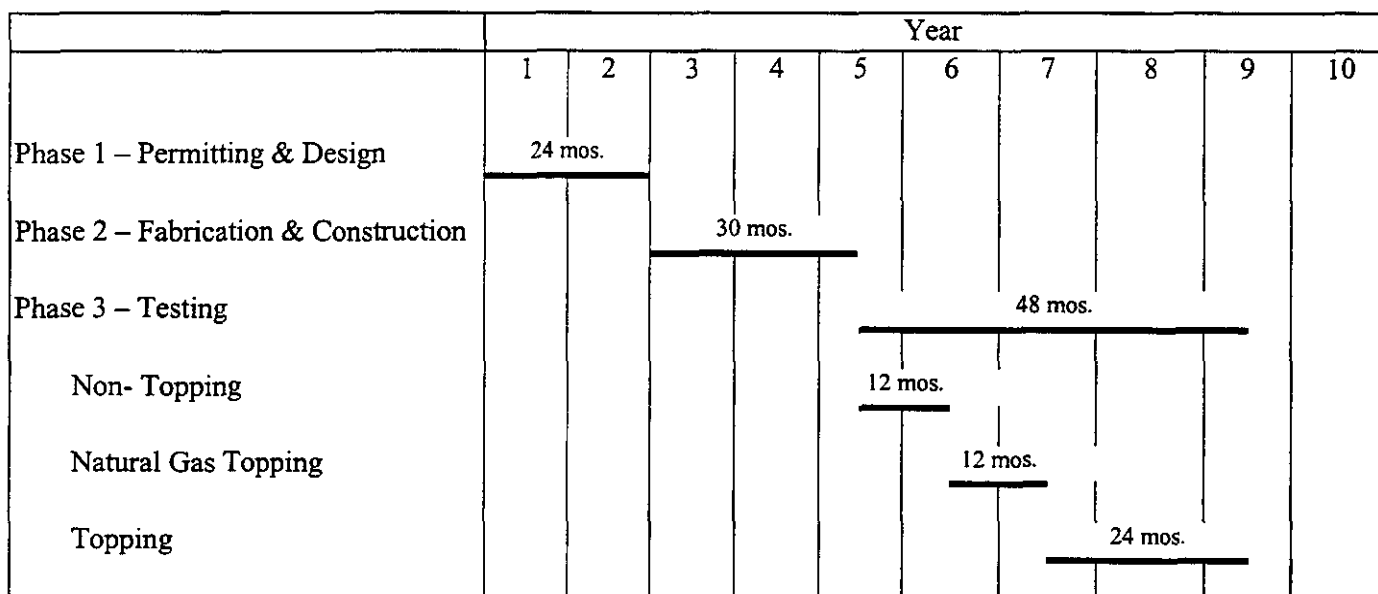


Figure 3 Lakeland Demonstration Plant Project Schedule

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LARGE-SCALE CFB COMBUSTION DEMONSTRATION PROJECT

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ABSTRACT

JEA's large-scale CFB demonstration project is described. Given the early stage of project development, the paper focuses on the design features being incorporated into the project, its role within the Department of Energy's Clean Coal Technology Demonstration Program, and the projected environmental performance. A description of the CFB combustion process is included.

I. INTRODUCTION

The U. S. Department of Energy (DOE) and the Jacksonville (FL) Electric Authority (JEA) have entered into an agreement that will preserve an important technology option in the federal government's Clean Coal Technology Program and provide the City of Jacksonville with an advanced, environmentally clean source of needed electricity in the 21st century.

DOE and JEA will share the costs of a \$435 million refurbishment of one unit of an existing municipal power plant with one of the world's most advanced technologies for burning coal. The project will boost the power plant's electricity production while reducing local air emissions and water consumption.

The new technology to be installed at the Northside Generation Station is called a "circulating fluidized bed combustor." The project will give Jacksonville the distinction of hosting the largest such combustor in the world — as well as one of the cleanest.

Circulating fluidized bed technology is an advanced method for burning coal and other fuels while removing pollutants inside the sophisticated combustor system. The combustor, itself, will eliminate more than 90% of pollutant-forming impurities. The JEA will add a further flue gas

scrubbing system to capture additional pollutants. Overall, more than 98% of the sulfur impurities in coal will be removed. The combustion technology is also flexible enough to burn other fuels including blends of coal and petroleum coke, a byproduct of oil refineries.

The JEA is the largest public power company in Florida and the eighth largest public power company in the United States. It currently serves over 335,000 customers and is experiencing an energy growth rate of more than 3% per year. The project is expected to increase the Northside Station electrical output by more than 2 ½ times.

For the DOE, the project will ensure that one of the most promising new coal technologies — attractive in both U.S. and international markets — remains in the nation's Clean Coal Technology portfolio.

Circulating fluidized bed technology has been installed in smaller, industrial-size plants but only recently has been considered for larger utility power plants. DOE helped test a 110-megawatt circulating fluidized bed combustor at a power station in Colorado in one of its earliest and most successful Clean Coal Technology projects. At nearly 300 megawatts — 265 megawatts of which will be supplied to the city electric grid (the rest is used to operate equipment at the power plant) — the Jacksonville project will more than double the size of the Colorado unit.

Foster Wheeler Energy Corporation of Livingston, New Jersey, will fabricate and install the new combustor. Previously, Foster Wheeler had planned to demonstrate the technology in a Clean Coal Technology project in York, Pennsylvania. When the local utility changed its power purchasing strategy, the power plant was no longer needed. The JEA's proposal offered the most favorable circumstances and assurances that the project would proceed.

DOE's originally-planned cost contribution will remain at \$74.7 million, with JEA providing the remainder, or nearly 83% of the total \$435 million costs. Unit 2 of the Northside Station, which has been out of service since 1983, will be outfitted — or "repowered" — with the new technology. Once the new combustion system is installed and brought on-line, the JEA plans to install a second identical circulating fluidized bed technology to repower Unit 1. The second unit would be privately financed. A third unit at the power station, a 564 megawatt oil-fired unit, will continue to operate.

Repowering Units 1 and 2 will result in lower particulates, sulfur dioxide and nitrogen oxide emissions than the Northside Station currently produces. Detailed engineering for the initial unit has begun and construction is expected to begin in August 1999. DOE's cost-sharing will include two years of demonstration test runs — April 2002 through March 2004 — during which both coal and coal-petroleum coke blends will likely be burned.

II. COMMUNITY COMMITMENT

An overarching consideration in the generation planning process is JEA's community commitment. JEA is committed to making Jacksonville "the premier city in the southeast in which to live and do business." In pursuit of this vision, JEA has established environmental performance standards for the project that exceed the current limits typically required for permits. The project will result in at least a 10% decrease in the emissions of sulfur dioxide, nitrogen oxides and particulate matter, along with at least a 10% decrease in consumption of ground water. This will be accomplished while generating over 2 ½ times the electrical output from the station as compared to the current operation. Therefore, JEA can help improve Jacksonville's environment while serving the growing economy and population.

Following the JEA Board's approval of the plan, the permitting process and detailed design have begun. A number of permits related to air quality, water use and water quality, solid waste handling and storage, etc. will need to be obtained. Prior to starting, and as the permitting process has progressed, JEA officials have been consulting with Northside and other Jacksonville residents, environmental interests, the business community and others to better understand their views, and get their input, and address concerns regarding the repowering plan.

III. PROJECT ORGANIZATIONAL STRUCTURE

The JEA has negotiated an agreement with Foster Wheeler that will address implementation of the extended boiler island scope of the project. This world-class showcase project will call on the services and expertise of two of Foster Wheeler's major operating groups: Foster Wheeler Energy Corporation, the world's leading supplier of CFB boilers, which will design and supply the new boilers for the Northside Station repowering; and Foster Wheeler USA Corporation (FWUSA), which will provide engineering, procurement, and construction management services on an open book basis for installation of the boilers and for furnish and erection of the air pollution control systems, chimney, limestone preparation system, and ash handling systems. Foster Wheeler Environmental Corporation, a subsidiary of Foster Wheeler USA, is providing environmental permitting services.

The remaining portions of the project will be implemented by JEA staff, which will be supplemented by Black & Veatch Corporation through a pre-existing alliance with JEA for engineering services. Construction and related services will be provided through other pre-existing alliances between JEA and Zachry Construction Company, Fluor-Daniel Inc., W.W. Gay Mechanical Contractor, Inc., and Williams Industrial Services Inc. This work will include an upgrade of the existing turbine hall; construction of the receiving and handling facilities for the fuel and reagent required to convert the plant from oil/gas firing to solid fuel firing; upgrade of the electrical switchyard facilities; and an ash management system.

IV. THE CLEAN COAL TECHNOLOGY PROGRAM

Under Public Law 99-190, the U.S. Congress provided authorization and funds to DOE to support the construction and operation of demonstration facilities selected for cost-shared financial assistance as part of DOE's Clean Coal Technology Program. In December 1985, Congress made funds available to DOE for conducting the first round of the Clean Coal Technology Program. In response to the solicitation, proposals were received and projects were selected by DOE for negotiation. In addition, a list of alternate candidates was established from which replacement could be made should any of the original selections not proceed. JEA's CFB combustor project has evolved from a project that was selected from the alternate list for demonstration. The overall Clean Coal Technology Program objective of the JEA project is to demonstrate the feasibility of scale-up of CFB technology to a size that will be attractive for large-scale utility operation.

The demonstration of JEA's CFB combustor project under the Clean Coal Technology Program will fulfill an existing programmatic need. Substantial deposits of coal exist as a fuel resource suitable for and capable of resolving critical energy issues; however, a number of obstacles exist that not only limit the general availability of coal but also act as a barrier to its increased use. These impediments include (1) concerns about environmental issues, such as acid deposition, global warming, polyaromatic hydrocarbon emissions, and solid waste; (2) availability of the technology; and (3) performance of the technology. Since the early 1970s DOE and its predecessor organizations have pursued research and development programs that contain long-term high-risk activities which support the development of innovative concepts for a wide variety of coal technologies through the proof-of-concept stage. However, the availability of a technology at the proof-of-concept stage is not sufficient to ensure its continued development and subsequent commercialization. Before any technology can seriously be considered for commercialization, it must be demonstrated. The risk associated with technology demonstration is, in general, too high for the private sector to assume in the absence of strong economic incentives or legal requirements. The congressionally-directed Clean Coal Technology Program provides a mechanism to accelerate the development of technology to meet the nation's near-term energy and environmental goals, to reduce technological risk to industry to an acceptable level, and to provide private sector incentives required for continued research and development aimed at finding solutions to long-range energy supply problems.

V. PROJECT SCOPE

The project will involve the construction and operation of two CFB combustors fueled by coal and petroleum coke to repower two existing steam turbines, each generating nearly 300 megawatts of electricity (MWe). The project will be located at JEA's existing Northside Generating Station, which currently consists of 3 heavy oil-and natural gas-fired steam units and 4 diesel oil-fired combustion turbine units. Units 1 & 3 are currently in operation and generate

steam from boilers that came online in November 1966 and June 1977, respectively. Unit 2 was completed in March 1972 but has not operated since 1983 due to boiler availability problems.

The new CFB will be capable of removing over 98% of the sulfur dioxide. However, to improve the overall economics and environmental performance, a polishing scrubber will be employed to minimize reagent consumption while firing petcoke containing up to 8.0% sulfur. The relatively low furnace operating temperature of about 1650°F would inherently result in appreciably lower nitrogen oxide emissions compared to conventional coal-fired power plants. However, the project will also include a new selective non-catalytic reduction system to further reduce emissions of nitrogen oxides. Over 99.8% of particulate emissions will be removed by a new baghouse or electrostatic precipitator.

In addition to the CFB combustor itself and the air pollution control systems, new equipment for the project will include a stack and fuel, limestone, and ash handling systems. The project will also require overhaul and/or modifications to existing systems such as the steam turbines, condensate and feedwater systems, circulating water systems, water treatment systems, plant electrical distribution systems, the switchyard, and the control systems.

New construction associated with the CFB combustor project will occupy approximately 75 acres of previously disturbed land at the Northside Generating Station. Solid fuel delivery will be accommodated by construction of new receiving, handling, and storage facilities. Limestone and ash storage and handling facilities also will be required. Wherever possible, existing facilities and infrastructure will be used for the project. These include the discharge system for cooling water, the wastewater treatment system, and the electric transmission lines and towers.

Project activities will include engineering and design, permitting, equipment procurement, construction, startup, and a 24-month demonstration of the commercial feasibility of the technology. Construction is scheduled to commence in August 1999 and finish in late 2001. Startup of Unit 2 will occur in early 2002, and demonstration of the technology will begin in April 2002. Startup of Unit 1 will occur in late 2002. During the two year demonstration, Unit 2 will be operated on several different types of coal and coal/fuel blends, to enhance the viability of the technology. Upon completion of the demonstration program for DOE, the facility will continue in commercial operation.

VI. CFB TECHNOLOGY DESCRIPTION

The CFB process offers the means for efficiently burning a wide variety of fuels while maintaining low emissions. Fuel is fed to the lower furnace where it is burned in an upward flow of combustion air. Fuel ash and unburned fuel carried out of the furnace are collected by a cyclone separator and returned to the lower furnace. Limestone is used as a sulfur sorbent, which

is also fed to the lower furnace. Furnace temperature is maintained in the range of 1500-1700°F by suitable heat absorbing surface. This process offers the following advantages:

- ☛ **Fuel Flexibility**
The relatively low furnace temperatures are below the ash softening temperature for nearly all fuels. As a result, the furnace design is independent of ash characteristics which allows a given furnace to handle a wide range of fuels.
- ☛ **Low SO₂ Emissions**
Limestone is an effective sulfur sorbent in the temperature range of 1500-1700°F. SO₂ removal efficiency of 95% and higher has been demonstrated along with good sorbent utilization.
- ☛ **Low NO_x Emissions**
Low furnace temperature plus staging of air feed to the furnace produce very low NO_x emissions.
- ☛ **High Combustion Efficiency**
The long solids residence time in the furnace resulting from the collection/recirculation of solids via the cyclone, plus the vigorous solids/gas contact in the furnace caused by the fluidization airflow, result in high combustion efficiency, even with difficult-to-burn fuels.

300 MWe CFB Boiler Design Features

Foster Wheeler's extensive experience in the areas of boiler and auxiliary equipment design, based on over 30 years of successful operating experience with large utility suspension-fired boilers, can be directly applied to the design of large CFB boilers.

Several aspects of CFB boiler design are unique to that technology, which include furnace sizing, cyclone design, ash cooler design, etc. These aspects, and Foster Wheeler's CFB designs and experience with them, are described in the following sections of this paper.

The 300 MWe CFB design described herein is based on a typical eastern bituminous coal and high sulfur petroleum coke. The following are the steam conditions and fuel analysis and on which the design will be based:

Steam Conditions at Turbine Throttle @MCR

Flow, SH/RH (pph)	1,993,591/1,773,263
Pressure, SH/RH (psia)	2500/547.7
Temperature, SH/RH (°F)	1000/1000

<u>Fuel Analysis:</u>	<u>Coal</u>			<u>Coke</u>		
	<u>Min.</u>	<u>Max.</u>	<u>Design</u>	<u>Min.</u>	<u>Max.</u>	<u>Design</u>
C, %wt	59.0	72.0	68.6	78.0	89.0	79.0
H	3.9	5.3	4.6	3.2	5.8	3.6
O	3.0	9.8	4.11	0.1	1.8	0.3
N	0.8	1.6	1.3	0.4	2.0	1.0
S	0.5	4.5	3.3	3.0	8.0	6.7
Ash	7.0	15.0	12.8	na	3.0	0.4
H₂O	na	13.0	5.2	na	15.0	9.0
Volatiles	30.0	36.0	35.6	7.0	14.0	9.0
HHV, btu/lb.	11,600	na	12,690	13,000	na	14,000

The design of the 300 MWe CFB is similar to earlier Foster Wheeler CFB designs and is shown in the attached figures. The boiler contains a single, water-cooled furnace. An INTREX (integrated recycle heat exchanger) receives the ash flow in the return leg from each cyclone, and contains intermediate and finishing superheater surface. Three steam-cooled cyclones are provided. The backpass is of parallel pass design and contains primary superheater, reheater, and economizer surface. A tubular air heater with flue gas inside the tubes follows the economizer in the gas path.

The process design for the 300 MWe CFB is the same as for other Foster Wheeler CFB boilers. Fuel and sorbent feed size, furnace velocity, furnace temperature, bed pressure drop, etc., are unchanged. Performance characteristics such as fuel burnout, sorbent utilization, boiler efficiency and emissions will be as good as, or better than, smaller boilers. Key design features include:

- Single furnace with division and wing walls
- Steam-cooled cyclones
- INTREX™ (integrated recycle heat exchanger)
- Parallel pass reheat control
- Start-up duct burners
- Water-cooled air plenum and fluidizing nozzles
- Fluidized ash cooler
- Fuel feed system
- Limestone preparation and feed system

Furnace

The main influences on CFB boiler configuration are the specified steam conditions and the fuel type. Compared with industrial boilers, the superheat and reheat duty of utility boilers is a greater percentage of the total input due to higher steam pressure and temperature. Higher feed water temperature in the utility boiler further increases the furnace heat duty due to a larger air heater duty which is transferred to the furnace.

The furnace temperature can be effectively controlled by changing the solids loading in the upper furnace by varying the primary/secondary air ratio and by changing the solids flow over the INTREX superheat surface.

The evaporative duty of the 300 MWe CFB unit is provided by the enclosure, division, and wing walls of the furnace. The furnace is a gas-tight enclosure formed from membrane tube panels cooled by natural circulation. Water-cooled partial division walls divide the furnace into three zones and so help evenly distribute gas and solids to the three cyclone separators. Six wing walls inside the furnace will provide additional evaporative surface. There will be no superheat or reheat surface located in the furnace. This arrangement of furnace and INTREX surface gives uniform heat removal and minimizes temperature variations.

To avoid erosion, a thin refractory lining is applied over metal studs in the lower furnace and around the openings to the cyclone. A patented tube arrangement at the top of the lower furnace refractory lining avoids local erosion. The critical dimensions of furnace height and depth have been controlled within Foster Wheeler's experience base to minimize scale-up risk.

Steam-Cooled Cyclone

The cyclone is one of the most important components in CFB boilers. Its efficiency is vital for the proper operation of the boiler. Proper cyclone efficiency will capture sufficient solids to ensure good bed quality which is manifested by proper furnace temperature and low temperature drop in the furnace, low carbon loss and low emissions.

The 300 MWe CFB boiler uses three steam cooled cyclones. Each cyclone is lined with refractory to protect against erosion. The refractory is Foster Wheelers standard one inch thick low cement refractory on studs. The stud density in the area of high solids impact is higher than in areas of lower erosion potential. Operating experience in all Foster Wheeler steam cooled cyclones shows that the refractory holds very well and is not a problem.

The scale-up of cyclones for large CFB boilers will run into two main limits: one is the possibility of lower collection efficiency and the other is that the cyclone will dictate the boiler height. Classic cyclone theory (which assumes no particle-particle interactions) predicts decreased efficiency with increased diameter. However, the inlet solids loading in CFB cyclones is so high that significant particle interaction does occur whereby a larger particle being collected carries other, smaller, particles with it to the cyclone wall. In fact, Foster Wheeler's experience is that larger CFB cyclones can have high efficiency.

Foster Wheeler invested heavily in a four-year test program (1990 - 1994) to optimize the cyclone capacity and collection efficiency. Over 90 different cyclone configurations and inlets were tested in various cyclone diameter test apparatus. The selected cyclone configuration is in operation in commercial CFB boilers, and performance has met or exceeded all expectations.

INTREX™ Heat Exchanger

In large CFB boilers, about 25% of total superheat (SH) duty is absorbed in the hot solids circulating loop, via in-furnace surface such as wingwalls or via surface in a heat exchanger such as an INTREX. As a boiler reaches utility sizes, Foster Wheeler uses the innovative and patented INTREX heat exchanger.

Solids returning from the cyclones flow into the inlet channels of the INTREX heat exchangers. During normal operation, the solids are passed into the SH cells by fluidizing both the inlet channels and the SH cells. During start-up the cells are bypassed by fluidizing only the inlet cells. By changing the mode of fluidization in the inlet channels and cells, solids flow to the cells can be controlled to change the superheat pickup in the INTREX, hence furnace temperature.

The INTREX enclosure is constructed from a steel shell with a thick refractory lining and comprises the inlet channel, superheat bundle cells, and a return channel to distribute solids evenly back to the furnace. The INTREX design for the 300 MWe CFB is based on the INTREX used at the NISCO plant. Five years of NISCO operating experience has shown that the design works very well. The INTREX provides the following advantages:

- **Furnace temperature control:**
As discussed above, the change of fluidization mode can effectively adjust the furnace temperature.
- **Reduced corrosion and erosion:**
High temperature SH surface located in the INTREX is not exposed to corrosive elements in the flue gas stream. This makes the INTREX excellent for firing corrosive fuels. Very low fluidization velocity in the SH cells (< 1.0 ft/s) and very fine particle sizes (~200 micron) eliminate the potential for erosion to the SH tubes.
- **Independent SH/RH steam temperature control:**
With all RH duty done in the backpass and most SH duty done in the INTREX, superheat and reheat steam temperature can be controlled independently over a wide range of conditions.

Parallel Pass Reheat Control

The backpass contains two parallel gas passes; the front pass houses the reheat surface and the rear, the primary superheater. The hot gas is biased by two gas pass dampers located underneath the HRA (Heat Recovery Area). Reheat temperature control is achieved without water spray by controlling the gas flow passing over the reheater, which causes no reduction in cycle efficiency compared with spray RH control. This design has been proven on Foster Wheeler utility boilers in sizes up to 930 MWe.

Start-Up Duct Burners

Duct burners, firing natural gas (with oil as backup) will be used for start-up to preheat the primary air stream which in turn uniformly preheats the bed material to the temperature needed

for solid fuel combustion. This preheating method maximizes the efficiency of bed preheating and so minimizes the amount of start-up fuel required, saving about 40% start-up fuel compared with start-up burners located on the furnace wall.

Water-Cooled Air Plenum and Fluidizing Nozzles

The air plenum under the grid at the base of the furnace distributes primary air to the fluidizing nozzles in the furnace floor. Foster Wheeler uses a water-cooled plenum, formed from tubing which then forms the furnace walls. The plenum is designed for high temperature gas so that boiler start-up time is minimized. Directional fluidizing nozzles are used to guide the bed ash flow towards the bed drains. These proven nozzle designs provide for low pressure drop, and minimizes the potential for back-sifting and pluggage.

Bottom Ash Cooler

The bottom ash cooler is required to maintain the desired furnace inventory and cool the ash to the temperature required by the bottom ash handling system. The 300 MWe CFB boiler uses the Foster Wheeler patented stripper/cooler design which is used normally for large CFB boilers or for high ash fuel in smaller boilers. The stripping (classifying)/cooling process consists of draining material from the bed and then, fluidizing this material in the stripper zone at a velocity sufficient to strip the required amount of fines from the stream and return these fines to the furnace. The balance of the material, which is primarily coarse, will pass through the next cooling zones to the ash drain in the floor of the last zone. These zones are fluidized and cooled by air from the air heater and from the primary fan. The stripper section is important for returning the fines (typically unburned carbon and unutilized limestone) to the furnace thereby increasing carbon burnup efficiency and reducing limestone consumption. The stripper/cooler also raises the boiler efficiency significantly by recovering the heat from bottom ash, as compared to cooling devices which do not recover this thermal energy.

Fuel Feed System

A total of twelve fuel feed points are provided; six in the front wall via gravity feed chutes and six in the rear wall via the solids return lines from the INTREX heat exchanger. This layout provides for uniform distribution of the fuel, which helps to maximize limestone utilization. From the day bins, fuel is extracted via gravimetric belt feeders, which carry the fuel to the front wall feed chutes and to drag chain conveyors and rear wall feed chutes.

The fuel feed system is designed to accommodate a positive pressure condition with the furnace balance point set at the cyclone inlets. Seal air is provided from the primary air fans to the belt feeders. These fans also provide air to the air swept fuel distributors. The air swept fuel distributor adds horizontal momentum to the fuel to assist in injecting it into the boiler. Seal legs of material are provided in the downspouts above the belt feeders. These legs are of sufficient height to seal against the maximum furnace pressures anticipated.

The proven air swept fuel distributors have been carefully designed to propel the fuel into the furnace in such a manner as to avoid hang-ups and back flow from the furnace and to distribute the fuel throughout the bed. They are the result of an extensive research program involving numerous flow models and operating experience. Air is admitted into each distributor at two locations in a carefully designed manner to maintain the proper velocity and flow pattern.

Limestone Preparation and Feed System

The raw limestone will be dried and ground to the appropriate size in rod mills. Three independent mill systems will be provided, one of which will be a spare system under most conditions of fuel quality. Prepared limestone will be pneumatically conveyed to day silos in the boiler area. From the day silos, limestone is extracted via rotary feeders and then pneumatically conveyed to the feed points. There will be three pneumatic feed systems installed, one of which will be an installed spare system. A total of twelve limestone feed points will be provided; eight in the front wall and four in the rear wall. This layout provides for uniform distribution of the sorbent, which helps to maximize limestone utilization.

VII. EMISSIONS PERFORMANCE

SO₂ emissions are easily controlled with limestone feed to the furnace. Over 90% SO₂ removal is typical and 95 to 98% removal is achieved in several units. The SO₂ removal requires no additional scrubbing equipment as the chemical reaction takes place in the furnace. The Ca/S ratio for 90 % SO₂ capture is normally around 2.0 for fuels with moderate to high sulfur content. SO₂ reduction is enhanced by good mixing in the bed and by increased excess O₂ level.

The sulfur dioxide emission limit of 0.15 lb./mmbtu is remarkable considering that the units will be designed to burn 100% petroleum coke containing up to 8.0% sulfur. A unique feature of this project is the integration of a semi-dry polishing scrubber into the CFB process. Most, if not all of the reagent for the polishing scrubber will be the high calcium oxide fly ash exiting the CFB. The result will be overall removal efficiencies greater than 98% at a lower calcium-to-sulfur ratio than with the CFB alone. Depending upon the ultimate fuel selection for the project, the reduced limestone consumption could possibly pay for the added capital cost of the polishing scrubber. However, another consideration in the decision to add the polishing scrubber was enhanced environmental performance in the area of hazardous air pollutant removal. While this benefit is impossible to quantify, JEA decided that the added capital cost was justified by furthering our commitment to improving the environment.

NO_x emissions are inherently low due to low furnace temperatures and staged combustion. Most of the NO_x is formed in the lower portion of the furnace, with NO_x emissions increasing with fuel volatile content, furnace temperature, O₂ level, and free lime available in the furnace and decreasing with an increase in the amount of char available. Therefore, minimizing excess O₂ in the furnace is important for NO_x control, which, however, is in conflict with SO₂ reduction. The

Foster Wheeler CFB process is optimized in such a way that the dense bed in the lower furnace provides a long residence time for char and limestone particles, thereby minimizing both SO₂ and NO_x emissions.

Additional NO_x control will be provided via an SNCR deNO_x system whereby aqueous ammonia is injected into the flue gas stream at the cyclone inlet. The cyclone provides for efficient mixing of the flue gas and ammonia and sufficient gas residence time at the optimum temperature for effective NO_x reduction. The combination of low NO_x at the furnace outlet with the additional reduction provided by the SNCR system will result in emissions below 0.09 lb/mmbtu.

Particulate control devices will be an integral part of the polishing scrubber process. Depending upon the final selection of scrubber technology, the particulate control will either be a pulse jet fabric filter or an electrostatic precipitator. The precipitator will only be allowed in conjunction with a circulating fluidized bed scrubber. Conventional spray dry absorbers will be provided with a fabric filter. Outlet emissions of particulate matter (total and size fraction less than 10 micron) will be less than 0.011 lb/mmbtu.

VIII. PROJECTED OPERATIONAL RESULTS

As noted previously, JEA strives hard to help both the business environment and the quality of life in the Jacksonville area. Towards that end, a requirement that JEA has imposed on the Northside Repowering Project is that the project result in a decrease in annual air emissions and groundwater consumption from the Northside Station.

Because Unit 2 has not operated since 1983, the baseline emissions from that unit are zero. Units 1 and 3 have been operating at annual capacity factors of less than 40%, firing either heavy oil or natural gas. Unit 3 will continue as a 564-MWe oil/gas-fired unit. With the exception of low NO_x (nitrogen oxide) burners on Unit 3, Units 1 and 3 are not currently equipped with emission control systems. As part of JEA's commitment to the local community in the implementation of this project, JEA has committed to a 10% reduction in the annual stack emissions of each of three criteria pollutants (i.e., sulfur dioxide, nitrogen oxides, and particulate matter) from the Northside Generating Station steam units (as compared to recent typical annual emissions). In achieving this objective, the combined emissions from the repowered Units 1 and 2 operating at annual capacity factors of 90% are expected to be less than or close to recent typical annual emissions from Unit 1 alone. This is particularly noteworthy considering that the electrical output of the combined Units 1 and 2 is expected to be almost five times that of Unit 1 in recent typical years.

Given that the emissions control systems will meet or exceed what is currently considered Best Available Control Technology, JEA anticipated that environmental permits would be obtainable. However, it was decided that maintaining emissions at current levels was not good enough, even with the significant increase in electrical output. As a result, JEA committed to going beyond the expected regulatory requirements and incorporated the otherwise unaffected Unit 3 into the

community commitment. Emissions from Unit 3 will be limited as required to achieve the station's overall 10% reduction target. It is expected that the goal will be met by operation of Unit 3 within the typical range of capacity factors and fuel blends that Unit 3 has experienced in recent years. However, if necessary to accomplish the goal, natural gas will be utilized to a greater extent or emission control systems will be added.

Groundwater consumption is an issue that JEA has focused much attention upon in recent years. Even with returning Unit 2 to operation, JEA has committed to reducing groundwater consumption by 10% from the 634,000 gpd that was consumed in 1996 (the most recent year at the time the project was approved). This represents a continuation of ongoing groundwater conservation initiatives, whereby consumption by the station has been steadily reduced from a high of 2.7 mgd in 1984 to the 1996 level of 0.634 mgd.

IX. CFB DEMONSTRATION PROGRAM

During the first two years of operation, through a series of operability, capability and performance tests, successful function of the unit will be demonstrated. Such operational testing will consist of three elements as follows:

Operability Tests - These tests will demonstrate boiler and associated equipment operability under various conditions. Operational and process parameters will be demonstrated through testing during:

- Start-up (cold and hot).
- Shutdown.
- Dispatch (maximum achievable loading rate from minimum load and maximum load reduction rate from maximum continuous rating (MCR) to minimum load).
- Operation at full load capability.

Reliability Determination Tests - This will involve collection of reliability data during the demonstration period to determine the overall reliability of the boiler and associated equipment. The data will be analyzed to determine the monthly and overall availability and capacity factor, plus it will identify the durations and causes of forced outages and forced load reductions.

Performance Tests - Performance guarantees will be verified through testing in accordance with applicable codes and EPA emission test methods. Specific testing will include or verify:

- Full load testing of the boiler at its MCR.
- Part load testing of the boiler at three part load conditions
- Power consumption of auxiliary equipment.
- Coals and coal/fuel blends, sorbent consumption and quality.
- Emission of NO_x, SO₂, CO, VOC, and Particulate.

- Emissions of trace elements and compounds.
- Boiler efficiency.

In addition to the above described operational testing, fuel flexibility tests will be conducted on individual coals and coal/fuel blends to evaluate boiler operability, capacity and performance. Such tests will be used to establish fuel, process parameters and boiler performance factors for use in determining the extent to which coal and coal/fuel blend characteristics can be varied.

During the operating period for each test, specific tests will be conducted to evaluate the following operating parameters:

- Fuel size and quality
- Sorbent size and quality
- Fuel and sorbent feed rates
- Combustion temperature
- Excess air
- Primary/secondary air split
- Fuel distribution
- Load following capability (rate of load change)

Factors related to boiler performance to be evaluated and reported on are:

- Crusher and fuel feed design
- Combustion and boiler sizing
- Distribution of heating surface
- Fans
- Particulate control device
- Ash handling system
- Ash disposal system
- Variation in operation economics for different potential fuel combinations

Conclusions would also be developed with respect to fuel, related operating and maintenance practices, costs, environmental compliance, and other factors. Total production related costs using the various fuels will be evaluated for use in future CFB installations.

Other significant aspects of the testing plan will be long term durability testing of the CFB system and off-line inspections to evaluate wear and fouling characteristics of the equipment.

X. CONCLUSION

Since its first commercial CFB boiler was commissioned in 1979, Foster Wheeler continues to be the world leading supplier of CFB boilers and offers the most experience in the CFB industry. CFB technology is now mature, and can be scaled up to utility boiler sizes of 300 MWe. The design of large CFB boilers in the 300 MWe size range is still very much experience-driven and is supported by Foster Wheeler's extensive experience with large suspension-fired utility boilers

and by Foster Wheeler's extensive CFB experience base. Foster Wheeler's CFB process and proven design features best meet the challenge of utility CFB boilers.

While CFB systems may be technically feasible, they are not commercially accepted at the 300MW scale at this time. The JEA Northside Station repowering project will become a major milestone in the further development and commercialization of state-of-the-art CFB systems at a scale that has not been demonstrated. The project will provide a domestic benchmark against which the U.S. utility, independent power, and financing industries can measure the scale-up ability of the CFB technology to the 300MW scale and accompanying economies of scale, as well as, the maintainability, availability, and reliability issues related to utility-scale CFB's.

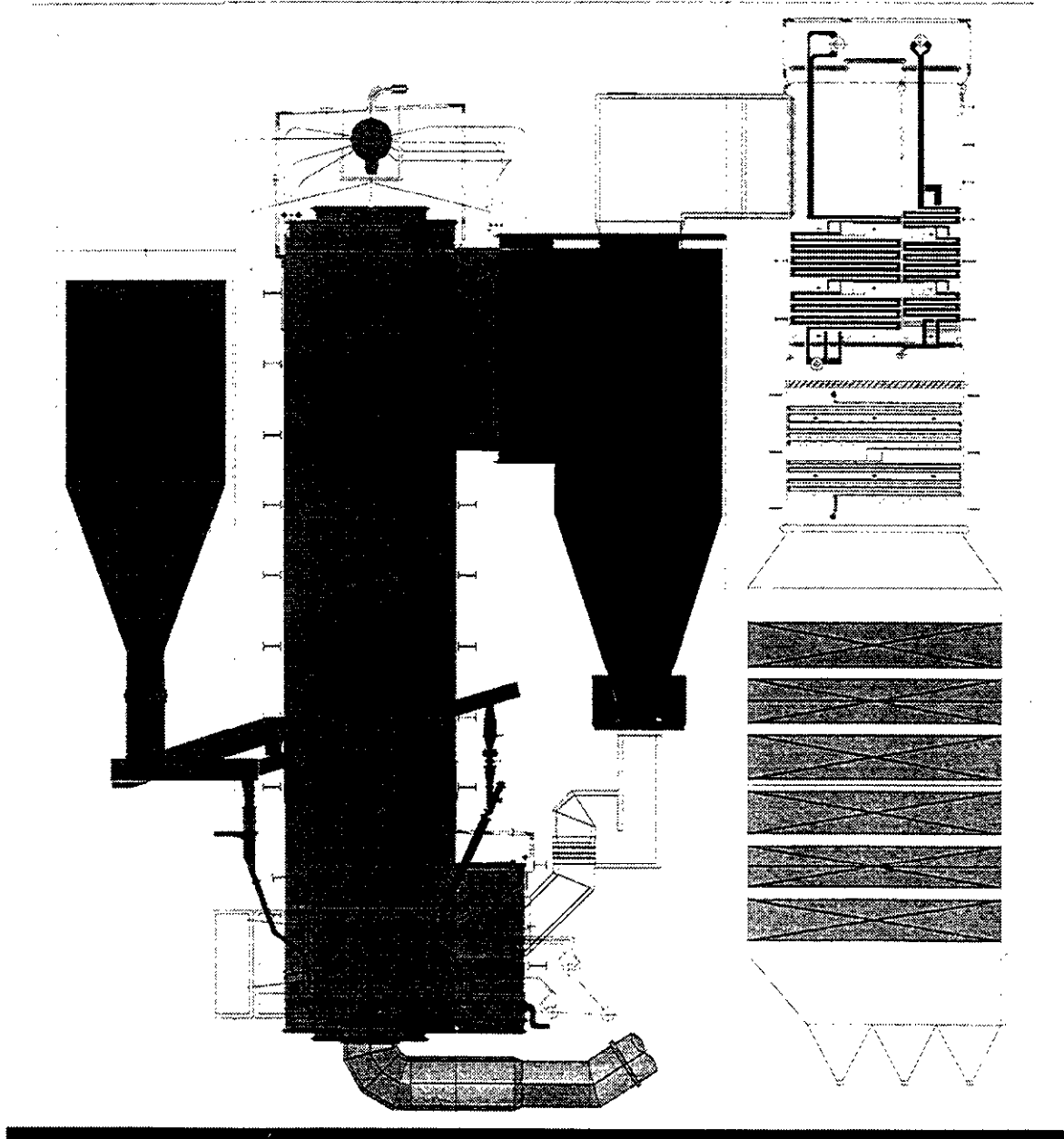
XI. REFERENCES

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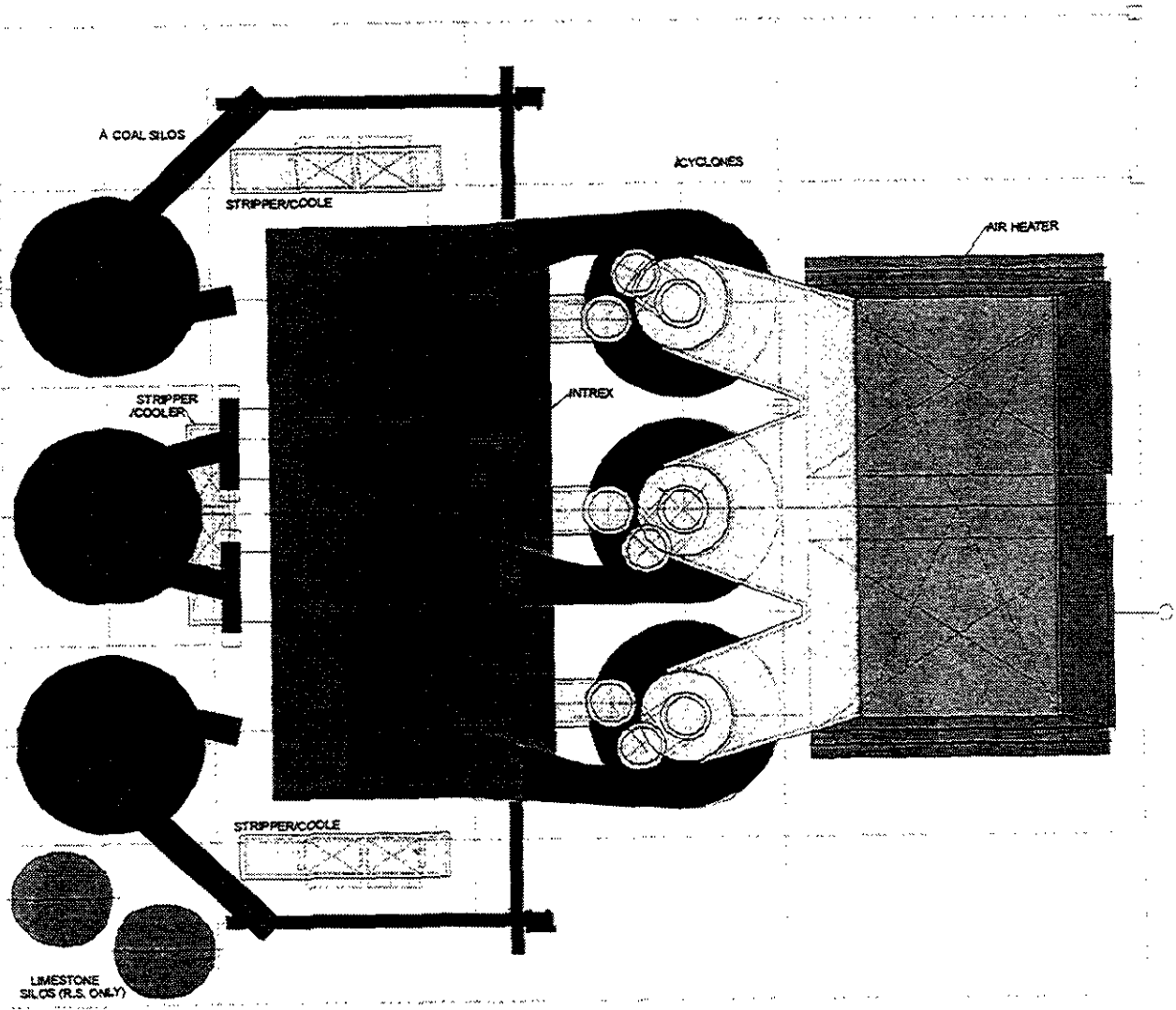
U.S. Department of Energy, "Notice of Intent to Prepare An Environmental Impact Statement and Notice of Floodplain and Wetlands Involvement for the Proposed Jacksonville Electric Authority (JEA) Circulating Fluidized Bed (CFB) Combustor Project", October 14, 1997

Foster Wheeler 300 MWe CFB



Side Elevation

Foster Wheeler 300 MWe CFB



Plan View

TECHNICAL SESSION III

Beyond 2010: Technology Opportunities
and R&D Needs

ADVANCED GAS TURBINES

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ABSTRACT

The status of gas turbine development is described, for three classes of gas turbines: Distributed Generation (0-50 MW), Intermediate Load Generation (On-peak and intermediate load power needs, 30-150 MW range), and Central Power Station Systems (large, base load combined cycles, >200 MW). Existing and potential Public-Private Partnerships are described. The status and content of the Next Generation Gas Turbine Program, which is in the discussion and planning stages at the U.S. Department of Energy, will be described. The relationship of these technologies to Clean Coal technology will also be described.

I. THE GAS TURBINE ASSOCIATION

The Gas Turbine Association (GTA) is a trade association with dues paying member companies supporting a paid staff in Washington, DC. Our website address is: <http://www.gasturbine.org/>
The mission of GTA is:

- Advocate government support of gas turbine technology development
- Advocate rational emission regulations
- Educate the government and general public regarding the economic and societal benefits of gas turbines

GTA member companies are as follows:

ABB Power Generation Inc.
AlliedSignal Power Systems Inc.
CAGT
Capstone Turbine Corporation
Catalytica Combustion Systems Inc.
Combustion Turbine & Combined Cycle Users Organization
Electric Power Research Institute
General Electric
Northern Engineering and Research Corporation
Parker Hannifin
Rolls-Royce Allison Engine Company
Sermatech
Siemens Westinghouse Power Corporation
South Carolina Institute for Energy Studies

Strategic Power Systems
United Technologies Corporation

GTA's overall recommendations for DOE's R&D Program were presented at the Department of Energy Strategic Visioning Workshop on Next Generation Gas Turbines in Austin, TX in February, 1999, see reference 1. The recommendations are as follows:

- 1) Continue ATS Program at accelerated funding levels for large (> 200 MW) and small (< 20 MW) gas turbines and combined cycles.
- 2) Develop of Flexible Gas Turbine Systems (On-peak and intermediate load needs, 30-150 MW)
- 3) Improve microturbine technology to increase efficiency for distributed power.
- 4) Support gas turbine/fuel cell hybrid system technology development.

Detailed recommendations for each program, including funding levels recommended for the FY2000 budget, were presented in testimony to the House Interior Appropriations Subcommittee on April 14, 1999. Since the founding of GTA in 1995, we have advocated appropriate R&D funding for the gas turbine industry. We believe this has had a positive impact in maintaining the ATS program so far, and expect it will be helpful in establishing the Next Generation Gas Turbine program.

II. GAS TURBINE TECHNOLOGY STATUS

Gas turbines for electric power generation can be considered to be in three classes, by application:

- Distributed Generation: 0 - 50 MW
 - ⇒ Small residential / commercial: 0-500 kW
 - ⇒ Medium commercial / industrial: 0.5-5 MW
 - ⇒ Large industrial: 5 - 50 MW
- Intermediate Load Generation: On-peak and intermediate load power needs
 - ⇒ 30 - 150 MW size range
- Central Power Station Systems: Large, base load combined cycles, > 200 MW

The status of these technologies is described in the following sections.

Microturbines (reference 2)

A typical Microturbine arrangement is a recuperated cycle with a centrifugal compressor and a radial inflow turbine:

The issues that will affect the success of Microturbines are:

- Technical Performance
 - ⇒ Efficiency = $f(\text{Recuperator Efficiency, Firing Temp., Parasitic Losses})$
 - ⇒ Reliability = $f(\text{Complexity, Durability, Ruggedness, Quality})$
 - ⇒ Emissions = $f(\text{Combustor Design, Fuel Composition})$
- Gas Booster Compressor (Cost and Reliability)

A summary of the Microturbine situation is:

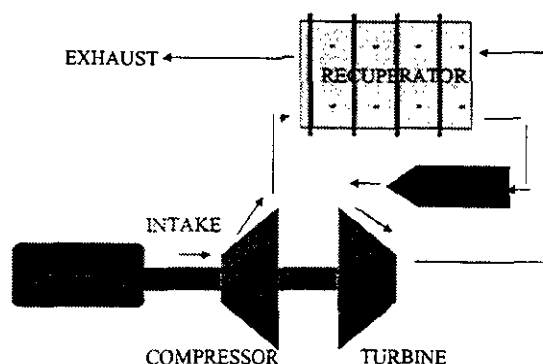
- Microturbines will be commercially available in 1999
- Products will vary by ruggedness, reliability, efficiency, emissions, cost, etc.
- Initial costs are likely to be closer to \$1000/kW **fully installed** than to \$350/kW due to high costs of inverters, recuperators, fuel boosters, and utility interconnection.
- Costs will come down with volume, but how fast remains to be seen.
- Microturbines must be foolproof, user-friendly, and marketed to volume buyers.

Studies and tests are underway for fuel cell / gas turbine hybrids. The status is:

- Best opportunity is in <5 to 30 MW size.
- Study by RR Allison of ~20 MW system
 - ⇒ 4 MW from gas turbine is typical.
 - ⇒ 71-72% efficiency estimated.
- Biggest issue for viability is cost of the fuel cell.
- Test program:
 - ⇒ 100 KW system being demonstrated
 - ⇒ 250 KW system being built
 - ⇒ 1 MW system being planned

Status of ATS Gas Turbines

TYPICAL MICROTURBINE ARRANGEMENT



The following points summarize the status of the small industrial and large utility units being developed in the ATS Program.

- Industrial, <20 MW
 - ⇒ Allison: Designed 701K (13 MW, 30:1, 2400 F, 40+%). Doing technology development using aero engines.
 - ⇒ Solar: Mercury 50 to ship 1Q99 (4.2 MW, recuperated, 4.2:1, 2125F, 40+%). Ceramic components in field test.
- Utility, >400 MW combined cycles (>200 MW gas turbines)
 - ⇒ GE: 9H to ship early '99, 7H to follow later (480/420 MW, 23:1, 2600F, 60%). Steam cooled rotor, stator.
 - ⇒ Siemens Westinghouse: 501G field test 2Q99 (430 MW, 27:1, 2725F, 60+%). Steam cooled transition piece. Working on "501ATS" model with steam cooled stator.

ATS: A Public / Private Partnership

The ATS Program is a good example of a public / private partnership:

- Public funding, cost shared with industry, is advancing the state of the art far further / more quickly than would have occurred otherwise.
 - ⇒ Sharing the risk makes it possible
- Public benefits are broad
 - ⇒ Reduced use of fuels
 - ⇒ Lower emissions
 - ⇒ Lowest cost of electricity
 - ⇒ Expanded jobs for the U.S. economy
 - ⇒ Universities developing people with gas turbine skills

This serves as a model for future DOE programs.

Intermediate Load Power Generation

There is currently no development underway of gas turbines intended for on-peak / intermediate load duty, 30-150 MW. They were not included in the ATS Program. This would be the subject of the Flexible Gas Turbine Systems (FGTS) Program. The needs to be met by this type of gas turbine are:

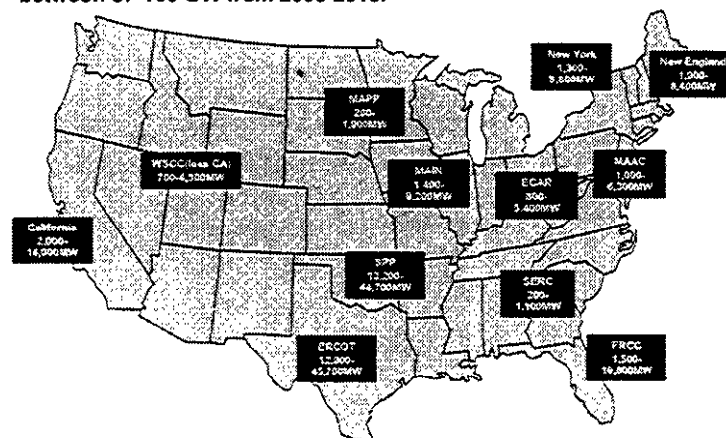
- Provide low cost of electricity compared to other gas turbines in cyclic, on-peak / intermediate duty.
 - ⇒ Avoid slow response time of frame type gas turbines
 - ⇒ Avoid high first cost (\$/kW) of current aeroderivatives

- ⇒ Provide higher efficiency than any simple cycle but without the high first cost of a combined cycle
- ⇒ Achieve low maintenance cost in cyclic duty.
- Enhance the operation and environmental performance of existing fossil fired steam plants.
- ⇒ Retrofit without the high cost of full repowering
- ⇒ Enable selling both on-peak and base load power
- ⇒ Short down time for maintenance
- Enhance renewable / synthetic fuel economics

The attributes of the product are:

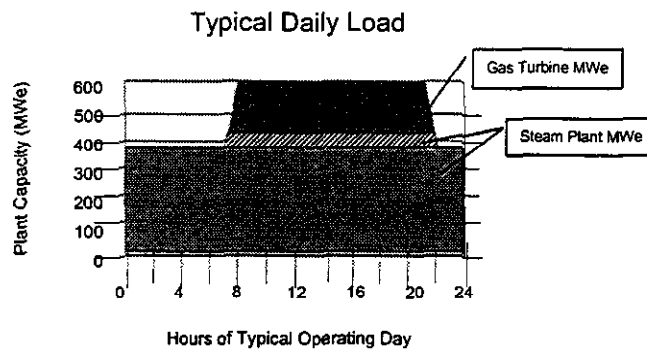
- 30-150 MW size range -- not covered by the ATS program
 - Designed for mid range power -- 500-5000 hours per year
 - ⇒ Efficiency in mid to high 40's percent
 - ⇒ Cost (\$/KW) closer to a large, frame type simple cycle than to a combined cycle
 - Designed for cyclic duty
 - ⇒ Cold start to full load in 10 minutes
 - ⇒ Able to take many full load start - stop cycles without significant reduction in useful life
- A market for FGTS gas turbines is to replace aging steam plants that were designed for base load and now are running in cyclic duty (ref. 3):

The market potential, based on displacement and load growth opportunities for AMGT technology in the U.S., is estimated to be between 37-160 GW from 2005-2015.



Another market for the FGTS is feedwater preheating of existing steam plants that are able to survive in the deregulated environment and can be enhanced to provide increased efficiency and the ability to sell both on-peak and base load power both on-peak and base load power. In the preceding figure, the gas turbine provides heat for feedwater preheating from its exhaust and intercooler. This allows more steam to be expanded through the steam turbine, increasing the output.

FEEDWATER PREHEATING PROVIDES BOTH ON-PEAK AND BASE LOAD POWER



FGTS Operating in Conjunction With a Vision 21 Plant: The DOE Vision 21 Program is intended to develop power plants that would take a variety of fossil fuels as feedstock, e.g. coal, natural gas, biomass. The products would be electricity and, depending on the specific market considerations of the plant, a mix of chemicals, syngas, transportation fuels and process heat. The V21 plants would operate base load.

FGTS gas turbines could operate in conjunction with V21 plants analogous to their operation in feedwater preheating service with base load coal fired steam plants. The FGTS gas turbine would produce on-peak electric power, using as fuel either natural gas or one of the synthetic fuels produced by the plant -- syngas or liquid fuels such as Methanol. When the gas turbine is operating, heat from the intercooler and turbine exhaust would be used in the process. This heat input would reduce the amount of feedstock needed to run the process. Thus the V21 plant would produce products both at base load and on-peak electric power.

Workshops have been held on this concept to get input from stakeholders, in March, 1997 and in October, 1998. Studies have been done or are currently underway or planned, funded by DOE, on feedwater preheating, cyclic duty issues, markets, public benefits, and operation with Energy Plex plants. The purpose of these studies is to define the attributes and benefits sufficiently, that decisions can be made on whether to proceed with development.

III. NEXT GENERATION GAS TURBINE PROGRAM

The Next Generation Gas Turbine Program is being considered for a public / private partnership. The topics under consideration are:

- Supporting Research and Technology
 - ⇒ Includes universities, national labs, manufacturers - same arrangement as in ATS.
 - ⇒ Continue working the many issues that remain in combustion, materials / coatings, design systems.
- Flexible Gas Turbine Systems
 - ⇒ FMGT has been identified as meeting the attributes needed and is one candidate system.
 - ⇒ Other attributes and systems may be identified.
- Vision 21 Applications
 - ⇒ Produce power and high value fuels / chemicals.
 - ⇒ Use coal, natural gas, biomass feedstocks.
 - ⇒ Minimize CO₂ by high efficiency, carbon capture.

The objectives of the program are:

- Market and U.S. Economy
 - ⇒ Reduce life cycle costs for the diverse set of power plants that will be added in a deregulated environment.
 - ⇒ Increase U.S. based suppliers' share in international markets.
- Public Benefits
 - ⇒ Reduction of emissions, e.g. CO₂ and NO_x for new and retrofit equipment.
 - ⇒ Increased electric system reliability.
 - ⇒ Increased choice of competitive generation options.
 - ⇒ Synergy with Vision 21 plants.
 - ⇒ Enabling technologies support other government missions, e.g. defense capability enhancement.

The proposed program elements and timing of the program are:

1. Supporting R&D, FY 2000-2006. Build on the successful ATS approach of R&D at universities, industry, and DOE facilities. Teams of FETC, National Laboratories, universities, industry and small business would work to solve technology issues. Technology areas to be addressed include:

- Advanced computing and testing for lean pre-mix, catalytic and fuel-flexible high pressure combustors,
- Materials and manufacturing development for alloys, coatings and ceramics,
- Heat transfer analysis and testing to develop novel cooling techniques, and
- Advanced computing and experimentation for improved aerodynamics.

2. System Definition and Conceptual Design, FY 2000-2001. For Flexible Gas Turbine Systems the tasks are:

Program definition and feasibility studies,
Evaluation of technical risks and development needs, and
Evaluation of markets and public benefits.

The results will be used to determine if DOE should go forward with partners to develop the systems.

3. Critical Technology Design and Testing, FY 2001-2005. The tasks are:

- Develop critical technology designs and perform sub-scale and full-scale component tests,
- Evaluate technical, economic, and environmental performance of the systems, and
- A decision point on whether to proceed with DOE supported engine and system integration tests.

4. Technology Transfer and Market Introduction, FY 2001-2007. The objectives of this program element are to ensure that the technologies and products developed are viable and marketable, and to maximize the public benefits and value derived from the program.

In order to determine if the program is progressing as planned, forums, workshops, and regional focus group meetings will be held; critical issues and progress will be discussed. Users' boards will be convened with DOE and other interested parties in an advisory capacity. Technology focus advisory councils will be convened in areas of combustion, materials, and aero / heat transfer.

5. Program Management Coordination: The objective of this program element is to ensure that the work of this program and gas turbine R&D funded by other organizations are complementary. A steering committee will be developed, consisting of DOE and other federal and state agencies to chart progress and direct the program. The program sponsors are envisioned to be DOE and other federal and state agencies.

IV. SUMMARY

The Next Generation Gas Turbine Program, if funded, will develop critical technologies that will enhance all types of gas turbines. Flexible Gas Turbine Systems will complete the spectrum of advanced gas turbine types to complement those developed by other programs. The results will improve costs, reliability, and environmental performance of electric power and will complement other government programs.

V. REFERENCES

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2. P. Bautista, "Micro Turbines: Opportunities, Issues, and R&D Challenges", presented at the Microturbine Technology Summit, Orlando, FL, Dec. 7-8, 1998.
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DEVELOPMENT OF CERAMIC ION TRANSPORT MEMBRANES FOR OXYGEN PRODUCTION

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DEVELOPMENT OF CERAMIC ION TRANSPORT MEMBRANES FOR OXYGEN PRODUCTION

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ABSTRACT

In partnership with the U.S. Department of Energy (DOE), an Air Products-led team (with Ceramtec, Eltron Research, McDermott Technology, NREC, Texaco, the Pennsylvania State University, and the University of Pennsylvania) is aggressively developing a new technology for air separation – Ion Transport Membrane Oxygen – based on the use of mixed-conducting ceramic membranes which have both electronic and oxygen ionic conductivity when operated at high temperature, typically 800 to 900°C. Under the influence of an oxygen partial-pressure driving force, the ITM Oxygen process achieves a high-purity, high-flux separation of oxygen from a compressed-air stream. By integrating the energy-rich, oxygen-depleted, non-permeate stream with a gas turbine system, the ITM Oxygen process becomes a co-producer of high-purity oxygen, power, and steam.

Under a recent CRADA entitled “Ion Transport Membranes (ITM) for Oxygen-Blown IGCC Systems and Indirect Coal Liquefaction,” Air Products and DOE quantified the potential benefits of an ITM Oxygen-integrated IGCC facility. Compared to the cryogenic oxygen base case, the ITM Oxygen technology can: improve thermal efficiency for the integrated IGCC system by about 3%; further decrease carbon dioxide and sulfur emissions; and, reduce the cost of generated electric power by more than 6%.

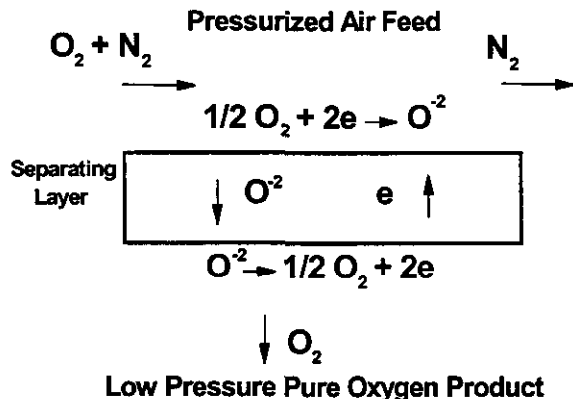
The ITM Oxygen development project will proceed in three phases. Phase I, which commenced under a DOE Cooperative Agreement in October 1998, is a 3-year effort focusing on construction of a technology development unit (TDU) for process concept validation tests at a capacity of 0.1 ton-per-day (TPD) oxygen. To accomplish this objective, the Air Products team will address relevant technical challenges in ITM Oxygen materials, engineering, membrane module development, and performance testing. During Phase I the team will also verify the economic prospects for integrating ITM Oxygen technology with IGCC and other advanced power generation systems. After at least one intermediate scale-up, Phase II & III activities will culminate with scale-up to a 25 to 50 TPD pre-commercial demonstration unit, fully integrated with a gas turbine. Meeting these challenges of developing cost-effective fabrication techniques for ITM Oxygen devices, and successfully integrating them with commercially available gas turbine engines, is key to bringing ITM Oxygen technology to the marketplace.

I. TECHNOLOGY DESCRIPTION

The ITM Oxygen process uses nonporous, mixed-conducting ceramic membranes, which have both electronic and oxygen ionic conductivity when operated at high temperature, typically 800 to 900 °C (Figure 1). The mixed conductors are complex formulations of inorganic mixed-metal oxides that are stoichiometrically deficient of oxygen, causing a distribution of oxygen vacancies in the lattice. Oxygen from the air feed adsorbs onto the surface of the membrane, where it dissociates and ionizes by electron transfer from the membrane. The oxygen anions fill vacancies in the lattice structure and diffuse through the membrane under an oxygen chemical-potential gradient, applied by maintaining a difference in oxygen partial pressure on opposite sides of the membrane. Meanwhile, an electronic countercurrent accompanies the oxygen anion transport. At the permeate surface of the membrane, the oxygen ions release their electrons, recombine, and desorb from the surface as molecules. Since no mechanism exists for transport of other species, the separation is 100% selective for oxygen, in the absence of leaks, cracks, flaws, or connected through-porosity in the membrane.

Figure 1

Oxygen Separation by Ion Transport



The solid-state diffusion of oxygen anions through mixed conductors is well-documented in the literature. The simplest approach follows that of Wagner. Assuming constant ionic conductivity that is much less than the electronic conductivity, the ionic flux can be expressed as:

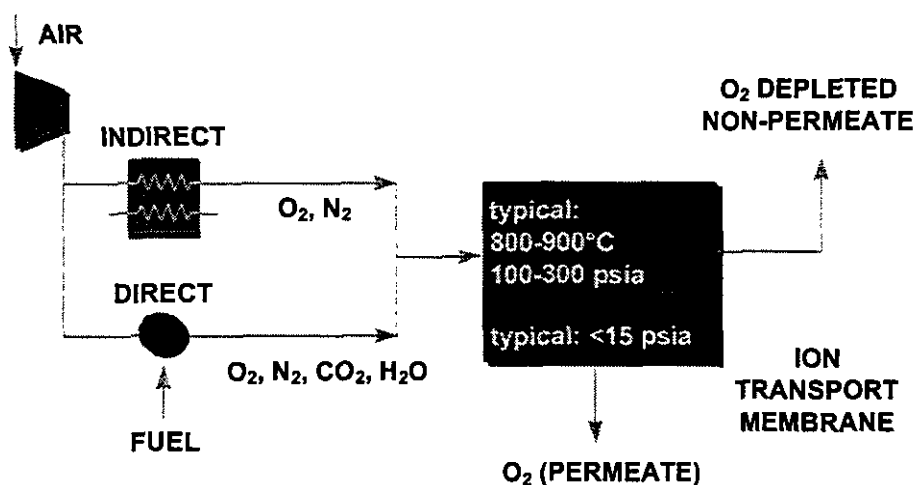
$$j_{O_2} = \frac{\sigma_i RT}{4Ln^2 F^2} \ln \left(\frac{P'_{O_2}}{P''_{O_2}} \right)$$

where j_{O_2} is the oxygen flux, F is Faraday's constant, L is the membrane thickness, n is the charge of the charge carrier ($=2$), R is the ideal gas constant, T is the absolute temperature,

P'_{O_2} is the oxygen partial pressure at the feed surface of the membrane, and P''_{O_2} is the oxygen partial pressure at the permeate surface of the membrane; σ_i is the ionic conductivity and is the only material property in the equation. This expression identifies the oxygen partial-pressure *ratio* as the driving force for the oxygen flux.

The inverse relationship between oxygen flux and membrane thickness identifies the need for thin-film ITM Oxygen structures, capable of supporting the pressure differential necessary to develop the oxygen partial-pressure driving force. To minimize the mechanical load imposed by the driving force, the process operating conditions constitute a medium-pressure air feed stream, typically 100 to 300 psia, and a low-pressure oxygen permeate stream, typically at a fraction of an atmosphere. Also, since the ionic transport mechanism is thermally activated, the process cycle must include the means to heat the pressurized air feed to high operating temperatures, either by indirect heat exchange or direct firing with an inexpensive fuel source (Figure 2).

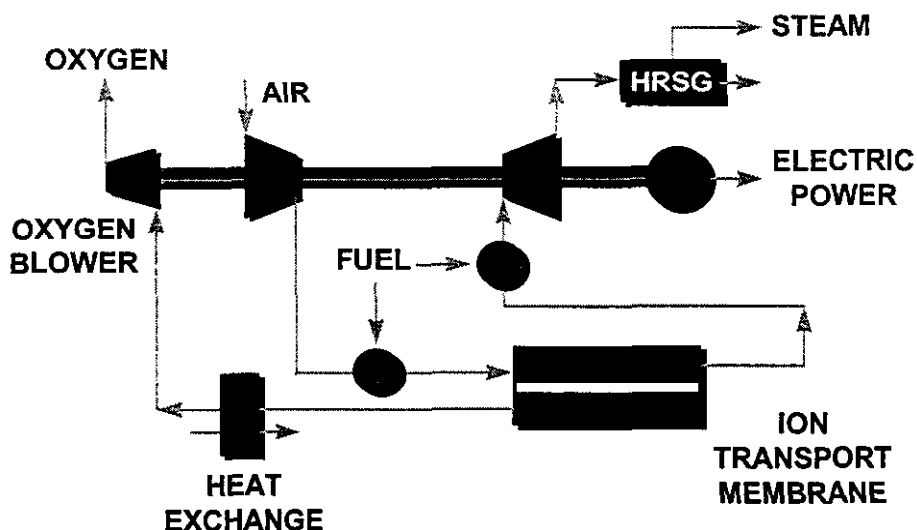
Figure 2 - Basic ITM Oxygen Process Components



To achieve acceptable cycle efficiency, the energy associated with the hot, pressurized non-permeate stream can be recovered by integrating the ITM Oxygen membrane with a gas turbine power generation system (Figure 3). This is accomplished by extracting air from the compressor discharge, withdrawing a portion of the oxygen by permeation through the membrane, and returning the partially-depleted non-permeate stream to the turbine inlet. The operating temperature of the ceramic membrane is above the compressor discharge temperature, but below the firing temperature characteristic of most (large) gas turbine engines. As a result, the cycle requires two separate combustors. The combustor upstream of the ITM Oxygen vessel raises the air stream temperature to the membrane operating value by burning a portion of the usual fuel input. A second combustor downstream of the ITM Oxygen vessel achieves the required turbine inlet temperature by burning the remainder of the fuel in an oxygen-lean atmosphere. The turbine exhaust may pass through a heat recovery steam generator to raise steam for export or combined-cycle power generation. The hot, low-pressure oxygen permeate stream is cooled and

compressed to the required use pressure. The full cycle, therefore, utilizes standard power cycle components and an inexpensive fuel, such as natural gas or synthesis gas from coal or another low-cost source, to co-produce oxygen, power, and potentially steam.

Figure 3 - ITM Oxygen Integration with Electric Power Generation



Integration with rotating equipment makes the ITM Oxygen process more capital-intensive, but significantly improves the overall efficiency. As a result, ITM Oxygen technology is generally appropriate for larger supplies of oxygen, nominally tonnage quantities, where it is projected to show significant cost benefits over tonnage cryogenic technology.

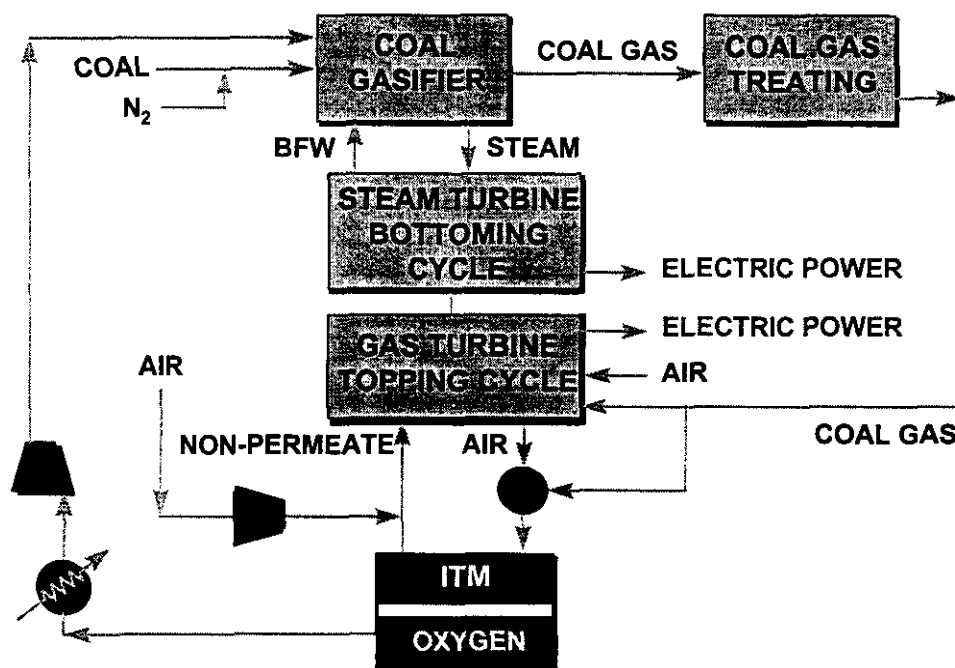
II. COMMERCIAL APPLICATIONS

ITM Oxygen technology is ideally suited to integration with power generation processes that require oxygen as a feedstock for combustion or gasification, or in any oxygen-based application with an export power market. As a result, the IGCC process is an ideal application for ITM Oxygen co-production technology.

Under a recent CRADA entitled "Ion Transport Membranes (ITM) for Oxygen-Blown IGCC Systems and Indirect Coal Liquefaction," Air Products and the U.S. DOE-FETC completed an initial quantification of the benefits of an ITM Oxygen IGCC facility. DOE-FETC provided the economic assessment for the base case, conventional, oxygen-blown IGCC facility — a partially (50%) integrated cryogenic air separation unit, along with a Westinghouse 501G gas turbine — and the costing model for the comparative economic analysis. Air Products provided the ITM Oxygen cost information. Electricity costs were calculated following EPRI TAG methodology (EPRI-4463-SR, 12/86).

Figure 4 depicts a flowsheet for the integration of an ITM Oxygen membrane into an IGCC facility.

Figure 4 - ITM Oxygen / IGCC Integration



The compressor section of a Westinghouse 501G delivers air at 294 psia, which is then heated to the ITM Oxygen operating temperature by direct combustion with a slipstream of clean coal gas. The membrane's chemical stability in high concentrations of carbon dioxide and water permits the use of this cost-effective, direct heating step. The high-purity oxygen permeate exits the membrane and is cooled prior to compression for use in the coal gasifier. The hot non-permeate stream emerging from the ITM Oxygen vessel is further heated by direct combustion with coal gas prior to introduction into the turbine section of the 501G. A supplemental air compressor adds sufficient air to replace the oxygen removed by this process cycle, maintaining the gas turbine near its peak power output to assure effective utilization of this key unit operation. The turbine exhaust then provides another heat source for the steam bottoming cycle.

A small, stand-alone cryogenic nitrogen plant cost-effectively generates the inert gas required for coal handling and conveying to the gasifier. This plant is approximately 35 times smaller than the cryogenic oxygen plant found in the base case.

Table 1 shows the cost and performance comparison for the cryogenic base case and the ITM Oxygen-integrated IGCC facility. The base case plant consumes 3,180 TPD of Illinois #6 coal

and 2,565 TPD of oxygen (95%), while producing a net power of 409 MW with a thermal efficiency (HHV) of 45.2%. The facility requires a Total Capital Investment (including on-sites, contingency, start-up, etc.) of \$641 million, resulting in an installed cost of 1,567 \$/kW. An EPRI TAG operating capacity factor of 65% and a coal price of \$30.60/ton were used for both economic assessments. Stated in terms of 10th-year levelized dollars, the cost of electricity was evaluated at 55.5 mills/kWh.

The ITM Oxygen-integrated plant consumes 3,176 TPD of Illinois #6 coal and 2,420 TPD of oxygen (99+%), while producing a net power of 420 MW with a thermal efficiency (HHV) of 46.5%. The facility requires a Total Capital Investment (including on-sites, contingency, start-up, etc.) of \$610 million, resulting in an installed cost of 1,453 \$/kW. Stated in terms of 10th-year levelized dollars, the cost of electricity was evaluated at 51.9 mills/kWh. The ITM Oxygen plant, including the supplemental air compressor, the additional combustor, ITM Oxygen modules, oxygen coolers, oxygen compressors, and the cryogenic nitrogen plant, saves 31% of the installed cost for air separation equipment.

Table 1 - Benefits of ITM Oxygen for the IGCC Application

	Cryogenic O ₂ Case	ITM O ₂ Case	Delta	% Change
IGCC Facility Capital Investment (\$Million)	641	610	(31)	(4.8)
IGCC Facility Capital Investment (\$/kW)	1567	1453	(114)	(7.3)
Power Production (MW)	409	420	11	2.7
Thermal Efficiency (%HHV)	45.2	46.5	1.3	2.9
Cost of Electricity (mills/kWh)	55.5	51.9	(3.6)	(6.5)

The economic benefits of ITM Oxygen technology for the IGCC application are excellent — a 2.9% improvement in thermal efficiency with a 6.5% decrease in the cost of generated electric power. The 11-MW improvement in power output equates to a 109 kWh/ton-O₂ decrease in specific power for oxygen production, approximately a 25% improvement for this application. The efficiency increase also produces a concomitant reduction in carbon dioxide and sulfur emissions. Therefore, integration of ITM Oxygen technology with IGCC offers the benefits of further improving system efficiency, resulting in better environmental performance and lower costs. Rarely do mature commodity products, such as oxygen and electric power, achieve such dramatic improvements.

The preliminary analysis undertaken in this study did not attempt to optimize integration opportunities. For example, the compressor section of the Westinghouse 501G gas turbine set may accommodate the small increase in air flow required to replace the oxygen removed by the ITM Oxygen modules. If so, the supplemental air compressor would not be required, and the savings on air separation equipment would improve to more than 40%, leading to further reductions in the cost of electricity.

The IGCC example teaches how ITM Oxygen technology may be effectively integrated with gas turbine engines. ITM Oxygen technology can also provide environmental and energy benefits to other industrial production technologies which require both oxygen and electric power, such as the cogeneration, pulp and paper, glass, steel, non-ferrous metals and chemical/refining industries. Furthermore, the cost improvements made possible by ITM Oxygen could enable growth in new applications of oxygen for efficiency improvements and emissions reductions in a variety of energy production and environmental cleanup technologies, for example, direct ironmaking processes and oxygen-enriched combustion technologies.

III. DEVELOPMENT PROJECT

The ITM Oxygen development project will proceed in three stages of development and scale-up toward commercialization. Phase I, which commenced under a DOE Cooperative Agreement in October 1998, is a 3-year, \$25 million effort, jointly funded by the DOE and Air Products, and focusing on construction of a 0.1 TPD technology development unit (TDU) for process concept validation tests. Phase I will also verify the economic prospects for integrating ITM Oxygen technology with IGCC and other advanced power generation systems.

To accomplish this objective, an Air Products-led team will address relevant technical challenges in ITM Oxygen materials, engineering, membrane module development, and performance testing. This technology development team comprises industrial and academic experts whose combined experience can meet all of these technical challenges: Ceramtec, a ceramics R&D and manufacturing company, will lead development activities in ITM Oxygen membrane fabrication and module assembly; Eltron Research will assist in developing optimal ITM Oxygen material compositions; McDermott Technology (formerly B&W Research) will provide its experience in high-temperature pressure vessel engineering, along with state-of-the art engineering capabilities in thermal and mechanical analysis for ceramic systems; NREC, a gas turbine and engineering development company, will provide know-how in gas turbine hardware design for the integration challenge with ITM Oxygen; and, Texaco will apply its extensive expertise in oxygen-blown gasification technology for IGCC and energy technology applications to help refine the process and economic analyses for ITM Oxygen-integrated IGCC facilities. In addition, ceramics experts from the Pennsylvania State University and the University of Pennsylvania will participate as consultants to the project, with Penn State also providing high-temperature materials testing capabilities. DOE's involvement will be overseen by FETC-Pittsburgh, through Dr. Arun C. Bose.

The Phase I project structure includes four principal development tasks. Task 1, Materials Development, involves an ITM Oxygen materials optimization effort incorporating performance and lifetime testing to define compositions that can meet the scale-up target design specifications. Air Products will select suitable materials with improved high temperature properties and ease of fabrication, using a statistically-designed property/composition database.

If appropriate, results from the other development tasks may be utilized to redefine the targets for key performance parameters.

In Task 2, Engineering Development, NREC will evaluate the critical technical and economic challenges associated with integrating the ITM Oxygen technology with a commercially-available gas turbine system. McDermott will perform thermal and mechanical analyses of membrane components and modules to facilitate their scale-up and improve performance. They will also complete a conceptual design and budgetary costing for a commercial ITM Oxygen process vessel to explore relevant design and safety aspects. Texaco will prepare case studies to compare the process economics of IGCC using cryogenic air separation and ITM Oxygen technology. Such economic assessments will help establish realistic performance targets for the other parts of the program. In addition, Air Products and the DOE will explore the most promising ITM Oxygen integrations with other advanced applications, which may include: coproduction of power, chemicals, and liquid fuels; pulverized coal combustion technologies; or, direct-reduction ironmaking technologies.

In Task 3, Membrane Module Development, Air Products and Ceramtec will utilize McDermott's structural analysis to refine the design of membrane components to minimize stresses. Ceramtec will scale up membrane components from the current sub-scale dimension to full size, using the optimum ITM Oxygen material compositions developed in Task 1. Ceramtec will also install a pilot ceramic membrane fabrication line to identify the critical factors in the fabrication process, which effect the yield of membrane components and to demonstrate the potential to attain economic targets for fabrication.

In Task 4, Performance Testing, Air Products will design and install a Technology Development Unit (TDU) comprising a prototype pressure vessel integrated with a pre-combustor to simulate accurately the commercial process conditions experienced by the ceramic membrane. This unit will help to demonstrate achievement of the target oxygen flux, while proving the compatibility of the membrane components with the commercially relevant process environment. Furthermore, it will demonstrate maintenance of high oxygen product purity and material stability in long-term tests at up to 0.1 TPD of oxygen production.

The overall goal of Phase I is to develop the ITM Oxygen technology to the point of demonstrating all necessary technical and economic requirements for further scale-up to the 5 TPD production scale. Phase II activities will focus on construction and testing of a 1 TPD unit, followed by further scale-up to 5 TPD oxygen production. The Phase II project will provide the design, engineering, cost, and scale-up data for a Phase III, 25 to 50 TPD, pre-commercial demonstration unit, fully integrated with a gas turbine and tested at a suitable field site. By the end of Phase III, sufficient demonstration of all essential aspects of the technology will have occurred to enable further industrial commercialization.

IV. SUMMARY

The ITM Oxygen process uses mixed-conducting ceramic membranes to produce high-purity oxygen at a high flux rate from a hot, compressed air stream. By integrating the energy-rich, oxygen-depleted, non-permeate stream with a gas turbine system, the ITM Oxygen process becomes a highly efficient co-producer of oxygen, power, and steam.

While Air Products and its ceramic fabrication partner, Ceramtec, have demonstrated the elements of ITM Oxygen technology with sub-scale, supported thin-film membrane components, the fabrication of membrane components must be scaled up to full-size units capable of withstanding the full operating conditions for extended time periods. In addition, various critical process integration issues must be adequately studied and understood to develop an economically viable ITM Oxygen process. These challenges represent significant technical hurdles that must be overcome before the technology reaches commercial viability, and they form the basis of the Air Products/DOE ITM Oxygen development project.

The development project will proceed in three phases of development and scale-up toward commercialization. Phase I, which commenced under a DOE Cooperative Agreement in October 1998, is a 3-year effort focusing on process concept validation tests, engineering development, and verification of the economic prospects for integrating ITM Oxygen technology with IGCC and other advanced power generation systems. After at least one intermediate scale-up, Phase II & III activities will culminate with scale-up to a 25 to 50 TPD pre-commercial demonstration unit, fully integrated with a gas turbine and tested at a suitable field site. Meeting these challenges of developing cost-effective fabrication techniques for ITM Oxygen devices, and successfully integrating them with commercially available gas turbine engines, is key to bringing ITM Oxygen technology to the marketplace.

V. ACKNOWLEDGMENTS

The authors gratefully acknowledge Steven Russek and Jianguo Xu of Air Products and Diane Revay Madden and Patrick Le of DOE-FETC for their leadership of the CRADA evaluation of ITM Oxygen integration with IGCC.

Clean Coal Technology in a Carbon Constrained World

presented at the
7th Clean Coal Technology Conference

June 22-24, 1999
Knoxville, Tennessee

by
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SFA Pacific, Inc.

Presentation Overview

Background

- SFA Pacific
- The Electric Utility Industry Is Changing
- Global Climate Agendas Heat up

CO₂ reduction options relative to coal & CCT

- Population
- Standard of Living
- Energy Efficiency
- Fuel Mix
- CO₂ Sequestering

Conclusions

SFA Pacific, Inc.

SFA Pacific Background

Basis of name: founded in 1980 as Synthetic Fuels Associates & does extensive work in the Pacific Basin

Perform technical, economic & market assessments for the private industrial sector

Principal work in coal technologies, residual oil upgrading & electric power generation

Niche is objective outside opinion and comparative analysis before companies make major investments

No vested interest in technologies, R&D or project development

SFA Pacific, Inc.

Representative SFA Pacific Clients

UTILITIES

Allegheny Power
CEA (Canada)
Electrabel
Electricite de France
EPDC
EPRI
Eskom
National Power
Ontario Hydro
Pacific Corp
Power Gen
RWE/Rheinbraun
So Ca Edison
Tokyo Electric Power
TransAlta

INDUSTRIALS

Air Products
ARCO/BP/Amoco
BHP
Chevron
Chinese Petroleum
Dow/Destec
DuPont/Conoco
Exxon/Mobil Oil
Idemitsu Kosan
Marathon Oil
Phillips Petroleum
Shell International
Statoil
Texaco
Weyerhaeuser

VENDORS

ABB
Babcock & Wilcox
Babcock Hitachi
Black & Veatch
Bechtel
Fluor Daniel
Foster Wheeler
GE
IHI
Kellogg/ Brown & Root
Kvaerner
Lurgi
MHI
Mitsui Engineering
Siemens/Westinghouse

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SFA Pacific Background in the CO₂ Issues

World Bank - Efficiency & environmental impact of coal use in China

United Nations - Several energy & environmental projects/conferences

China - International coal consultant to the People's Republic of China
National Response Strategy for Global Climate Change

Global Environmental Facility (GEF) - Recommendations & suggestions on coal technologies in a carbon constrained world

US DOE

- Review of policy & energy technology sections of 1995 IPCC draft
- Analysis of commercial & advanced CO₂ control options for electric power generation (see my paper presented at the GHGT-4 Conference at Interlaken, Switzerland August 1998)

Several private client analyses

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The Electric Utility Industry Is Changing

The great Chinese curse: "May you live in interesting times"

- Deregulation, increased competition & globalization
- Forcing short-term economic focus

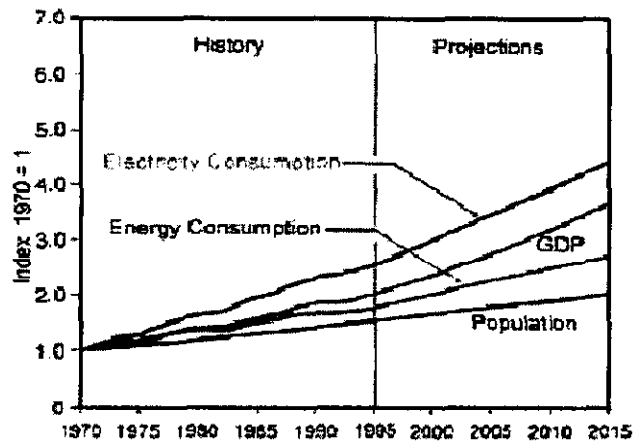
Uncertain future for coal-based power generation:

- Existing low-cost coal-fired power plants face more severe emission limitations on NO_x, SO₂, HAP, PM_{2.5}, & solid wastes
- Coal is not competitive with NGCC for new capacity at NG prices below about \$4 per MM Btu (current industrial NG prices are \$2-3)
- Future coal-based power generation may be required to meet the same emission limits as NGCC on a emission per MWe basis
- CO₂ could be the "mother of all emission limitations" for coal

Coal requires innovative CCT options to survive

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World Energy, GDP & Population Trends Clearly Show Electricity is the Energy of the Future



Source: 1997 US DOE/EIA International Energy Outlook

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Deregulation Changes Yet to Be Resolved

Positioning your company for success in a deregulated market

- IPP subsidiaries, power marketers & convergence of power & gas
- Restructuring: takeovers, asset resales & upgrades to create large, more economically efficient Genco's, Transco's & Disco's
- ISO of transmission systems (Transco's) to assure system reliability
- Time of day rates (both buying & selling) to improve system utilization

Existing coal units are the likely the big winner of deregulation

- Lowest marginal dispatch cost & great potential for upgrading

Emissions laws will set coal vs NG economics in the future

- CCT help reduce the cost of add-on NO_x, SO₂, PM_{2.5}, & HAP reduction
- CO₂ reduction does not kill coal thanks to CCT with CO₂ sequestration

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Global Climate Agendas Heat up

Totally irrelevant if there really is global climate change due to fossil fuels because it has become a political issue

Kyoto protocol has several fatal flaws

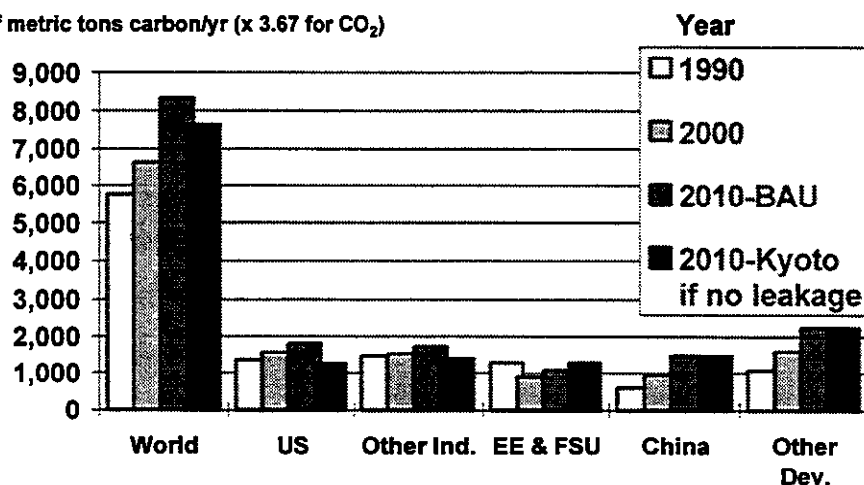
- Currently ignores developing nations which represents most of the entire world growth in CO₂
- "Leakage" would likely increase CO₂ growth by economically forcing CO₂ intensive industries in Annex I nations to move production to cheaper but less efficient & coal-based developing nations like China
- Most of the reduction burden on US - 26% reduction from year 2000
 - Highly unlikely that the US will ratify - "It's the economy, dummy"
 - See the October 1998 EIA Kyoto Report to Congress - SR/OIAF/98-03

Must begin to think about the better protocol after Kyoto fails

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World Carbon Production From Fossil Fuels By Region

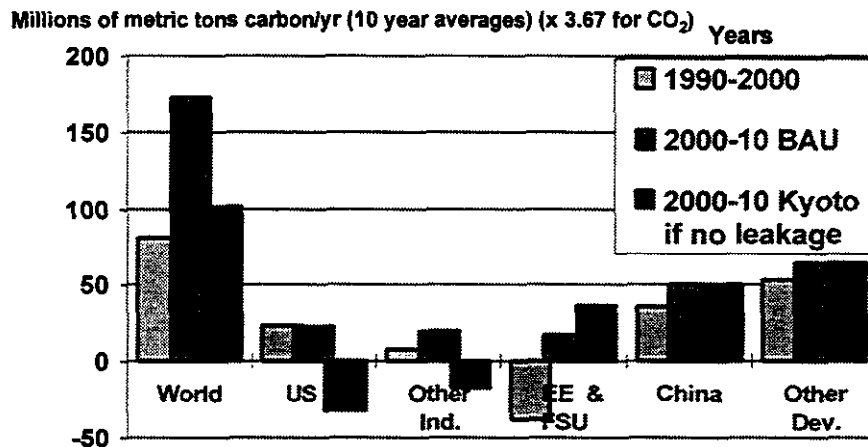
Million of metric tons carbon/yr (x 3.67 for CO₂)



Source: SFA Pacific from 1998 US DOE/EIA International Energy Outlook

SFA Pacific, Inc.

World Carbon Annual Growth Rate From Fossil Fuels by Region



Source: SFA Pacific from 1997 US DOE/EIA International Energy Outlook

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Power Generation Will Be Forced to Meet a Disproportionate Share of Any CO₂ Reductions

Environmental hypocrites driving urban assault vehicles (SUVs) have more votes than CO₂ intensive industries

Power plants can not move to China, as many CO₂ intensive industries in Annex I nations will, if faced with carbon taxes

Large potential for efficiency improvements in power gen.

- Uprate existing power plants

Large potential to replace coal with biomass & natural gas

- Biomass cofiring & NGCC (new, repowering & cogeneration)

Large point sources of power gen. reduce CO₂ recovery costs

- Lowest \$/CO₂ avoided costs are existing coal plant modifications

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CO₂ Reduction Options

CO₂ emissions & growth rate are simple to calculate:

people x GDP/person x energy/unit GDP x CO₂/unit energy

Only control options are:

- **Population** (number of people)
- **Standard of living** (GDP/person)
- **Energy efficiency** (energy/unit of GDP)
- **Fuel mix or CO₂ sequestering** (CO₂ /unit energy)

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Reducing the Developing World's Population Bomb

Lowest cost option

- **However, emotional, political & religious overtones**

Nevertheless, it is well known that populations stabilize

- **Primarily as standard of living increases to just a minimum level**
- **Secondarily as woman are allowed the same rights & education as men**
- **Italy & Iran are prime examples**

Electrification is essential to obtain the minimum standard of living required to stabilize population growth

- **Even if this involves construction of coal-fired power plants**

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Standard of Living

Essential to increase standard of living in the developing nations as quickly as possible to stabilize world population

Reducing standard of living with a recession works; however, only effective long-term for rich industrialized nations

- Well demonstrated by the 1973 & 1979 oil price shocks
- Carbon taxes will have same effect as oil price shocks
- EIA's Oct. 98 Kyoto Report to Congress - \$348 per ton carbon tax
- Carbon tax of \$10/MM Btu coal or \$40/bbl crude oil is the "lose-lose" approach, as this increases energy costs & reduces economic growth
- Recessions or drop in standard of living are not politically popular nor good for non-government funded workers

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Energy Efficiency

This is the "win-win" approach as this reduces energy end-use costs & increases economic growth

Two fundamentally different approaches to higher efficiency

- Advanced central power plants (favored by regulated utilities)
- Cogeneration (favored by deregulation & distributed generation)

Upgrade existing coal power plants

- Upgrade steam cycle
- Natural gas repowering (hot windbox allows continued coal use)

Once coal is competitive again for new capacity, requires CCT with the maximum flexibility for an uncertain future

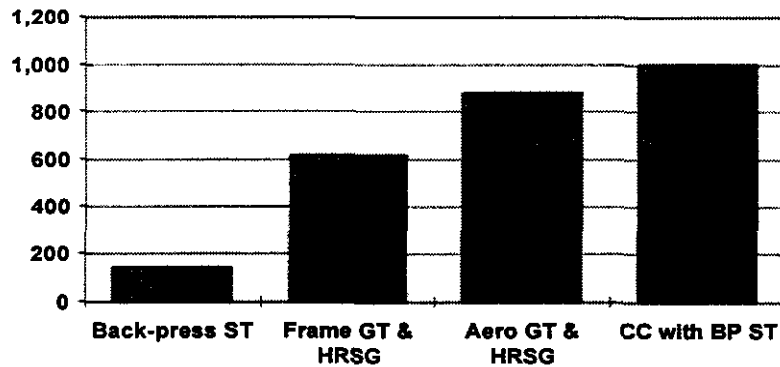
- Favors gasification due CO₂ recovery potential plus efficiency & flexibility of polygeneration (cogeneration steam & power + syngas)

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Maximum Power in Total Cogeneration Clearly Favors Gas Turbines Over Steam Turbines

For a given cogen heat host, 5-8 times more power with GT vs ST

Power-to-Steam ratio: kWe per t/hr cogen 10 bar steam (no steam to condenser)



Source: SFA Pacific from GE data

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Large Cogeneration Potential for CCT

Cogeneration can develop where ever new baseload power is needed & electric power generation is deregulated

- Favors gasification due to the high power to cogen heat ratio of GT plus the simpler & cheaper quench designs used in polygeneration
- Gasification of low cost petcoke & residual oil for refinery polygeneration already happening since deregulation in Europe & US

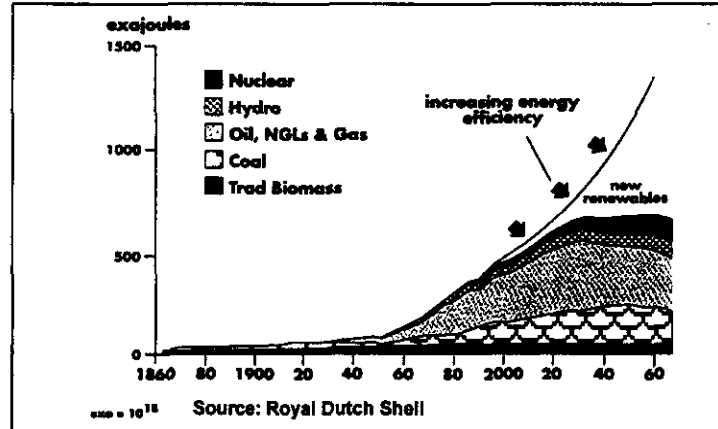
Major cogen opportunities in China's massive & growing coal markets, however, only after electric power is deregulated

- About 60% of China's current coal use is industrial, only 40% utility
- China industrials already operates over 5,000 MW_{th} of "world class" Texaco & Shell gasification for syngas (ammonia fertilizer) production
- Potential for massive coal gasification based-polygeneration if this 70% + efficient (fuel charged to power) could be sold to the grid

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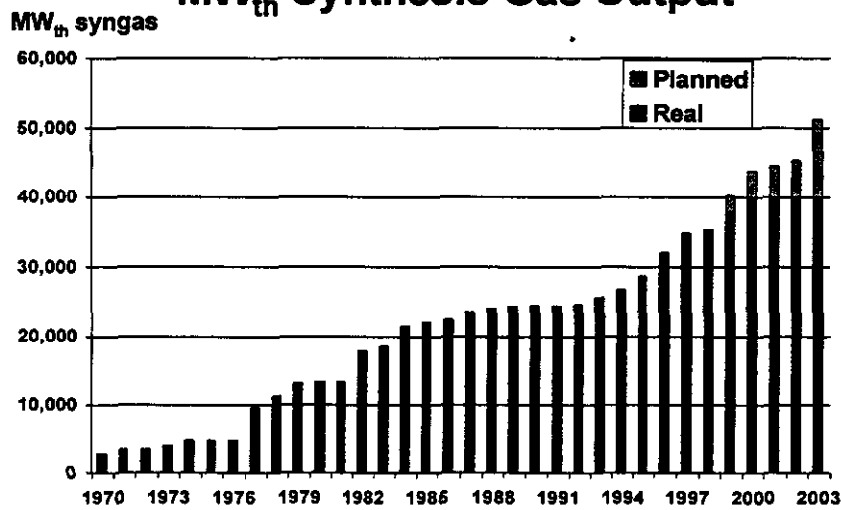
World Energy Supply Projection

Massive Expensive Renewables or Increased Energy Efficiency?
Most of World Coal Growth Is For Power Generation in China
Cogeneration via Gasification Could Double That Efficiency



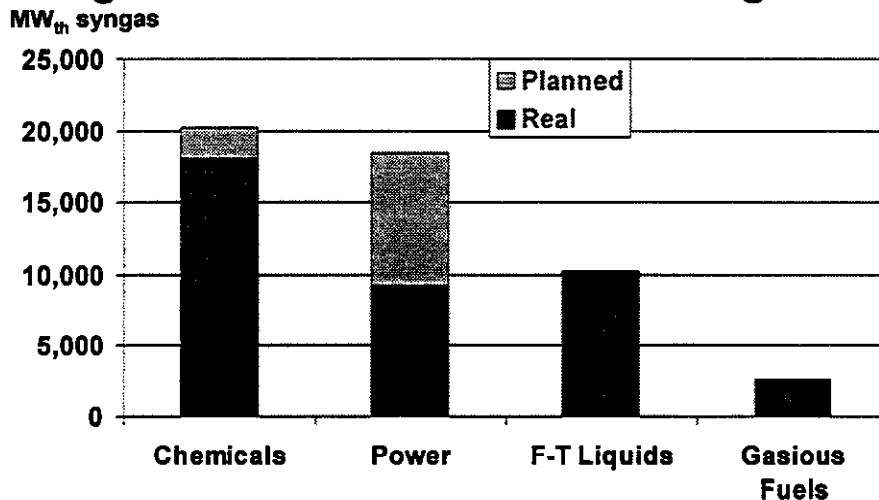
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Cumulative Worldwide Gasification Capacity in MW_{th} Synthesis Gas Output



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Gasification by Application Large Growth in Power due to Deregulation



Source: SFA Pacific, Inc for the US Department of Energy

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Fuel Mix Changes Hurts Coal

Natural gas to replace coal is the big winner of CO₂ reductions

Also favors life extension of existing nuclear plants & co-firing biomass in existing coal boilers where ever economical

Nuclear, renewable, reforestation & biomass are oversold

- New nuclear is non-competitive & cogen limited by steam cycle
- Wind-turbines suffer from low annual capacity & need for back-up
- Requires about 3 square miles of reforestation @ 2 tons C per ha/yr (until full grown) for one MWe of coal-based power to be CO₂ neutral
- Biomass is not cost effective in nations with high land & labor costs
 - Requires >\$250/acre/yr (\$618/ha/yr) for US farmers to consider growing biomass; @ 5 tons C per ha/yr, the C avoidance cost is >\$124/ ton

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CO₂ Sequestering Saves Coal

Transforming the debate since CCT with CO₂ sequestering is usually more economical than renewables

Requires a large CO₂ point source, high purity recovery, compression to high pressure (about 80% of the total costs)

Critical to test large CO₂ sequestering options (20% of costs)

- Statoil currently doing a 1 MM mt/yr CO₂ sequestering test via injection in a deep saline aquifer under the North Sea

CO₂ sequestering options for coal

- CO₂ recovery from existing gasification plants
- Flue gas scrubber or maybe O₂ combustion from existing coal power plants; however, major reduction in both net capacity & efficiency
- New gasification with CO₂ recovery & H₂ for repowering central power plants or industrial polygeneration

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Proposed CO₂ Sequestration Projects That Reduce Costs Via Slight Byproduct Value

Dakota Gasification - CO₂ recovery from existing coal gasification SNG for enhanced oil recovery (EOR) in Canada

China - CO₂ recovery from existing coal gasification ammonia for deep un-mineable coal bed methane (CBM) recovery

Alberta, Canada - CO₂ recovery from existing coal boilers or other sources for improved tar sands EOR

Shell Oil Pernis Refinery - CO₂ recovery from existing residual oil gasification polygeneration for use in greenhouses

BP/Amoco & ARCO Alaska North Slope - CO₂ recovery from existing gas turbines or others sources for improved EOR

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Conclusions

Current Kyoto protocol would make world CO₂ growth worse

- Ignores the population & energy growth of developing nations
- If faced with carbon taxes, many CO₂ intensive industries in Annex 1 nations could be economically forced to move production to China

Electric power generation will be forced to meet a disproportionate share of any CO₂ reductions

- Environmental hypocrites driving SUVs have more votes
- Can not move electric generation to China
- Large point sources of power plants lower CO₂ avoidance costs

New capacity additions will favor high efficiency technology

- Deregulation & the use of gas turbines clearly favor cogeneration
- Gasification based polygeneration is already happening in refineries

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Conclusions

Electric power generators of the future must objectively assess all options in the light of CO₂ politics & deregulation

- Reforestation & cofiring biomass is quite useful but limited
- Existing nuclear life extension & existing coal & nuclear uprating
- Natural gas repowering of existing coal units via hot windbox
- High efficiency of GT based cogeneration via NG & coal gasification
- However, coal's long-term survival in a carbon constrained world require CO₂ sequestration which clearly favor coal gasification

Keys to honest worldwide CO₂ reduction with CCT

- US energy policy reforms that encourage CO₂ reduction by "win-win" efficiency & CO₂ sequestering incentives, not "lose-lose" CO₂ taxes
- China energy policy reforms that allow 75% efficient power from coal gasification based polygeneration to be sold to the grid at a fair price

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An Advanced Coal Gasification
Power Plant With CO₂ Control Capability

Dwain F. Spencer
Principal, SIMTECHE

Introduction

Over the last two years, SIMTECHE, under contract from the U.S. DOE-FE, has been conducting engineering and economic evaluations of various Advanced Integrated Coal Gasification Combined Cycle (IGCC) Power Plants, including the implications of 90% CO₂ control capability. The most recent evaluation has focused on a CO₂ hydrate slurry, separating the product gas, primarily hydrogen, and regenerating a high pressure CO₂ stream, plus some H₂S, in a flash reactor. This high pressure stream can then be further compressed or upgraded, if necessary, to meet CO₂ sequestration or utilization requirements.

The power plant design utilized includes a) a Texaco gasifier, at high pressure, to generate the syngas, b) a radiant syngas cooler, only to recover much of the sensible heat from the gasifier, c) a two stage catalytic shift reactor to produce a predominantly H₂-CO₂ rich gas, d) the SIMTECHE CO₂ hydrate separation process for CO₂ removal, e) additional product gas cleaning, i.e. H₂S removal and sulfur recovery and f) the use of an advanced hydrogen fueled power cycle proposed by Westinghouse/Mitsubishi for power generation.

This paper will provide overall heat and material balances for the process, plant gross and net power output, and an estimate of overall power plant efficiency and costs of carbon control. The work is still in progress, so the final performance and more definitive cost estimates will be provided in a final report to DOE. This advanced power cycle is one type of Vision 21 plant and can benefit from many of the technology developments emanating from that program.

The analysis will also consider a two potential, terrestrial sequestration options with their gaseous CO₂ injection requirements and preliminary sequestration cost estimates. This analysis will provide a range of performance and costs for CO₂ capture, regeneration, and sequestration from an advanced IGCC plant, and demonstrate the potential for a hydrogen based fuel system for power generation from coal.

Process Description

Overall Powerplant

An overall powerplant process flow diagram is shown in Figure 1. The basic IGCC powerplant design has the following features:

- * A high pressure (750 psia) Texaco Partial Oxidation Gasifier with Radiant Syngas Cooler (SGC) only, followed by a water quench
- * Texaco proprietary carbon scrubbers for particulate removal
- * Cryogenic Air Separation Units to provide oxygen for gasification and combustion
- * Two stage sour shift for 95% conversion of the CO to CO₂
- * Gas Cooling to CO₂ hydrate formation conditions
- * The SIMTECHE CO₂ hydrate separation process for removing CO₂ from the product stream and regenerating a high pressure CO₂ stream
- * CO₂ compression, pipelining and injection for sequestration
- * Product gas (primarily H₂) acid gas (H₂S) removal and elemental sulfur recovery; reheat of the product gas
- * Utilization of the Bannister power cycle for power generation.

Plant Discription

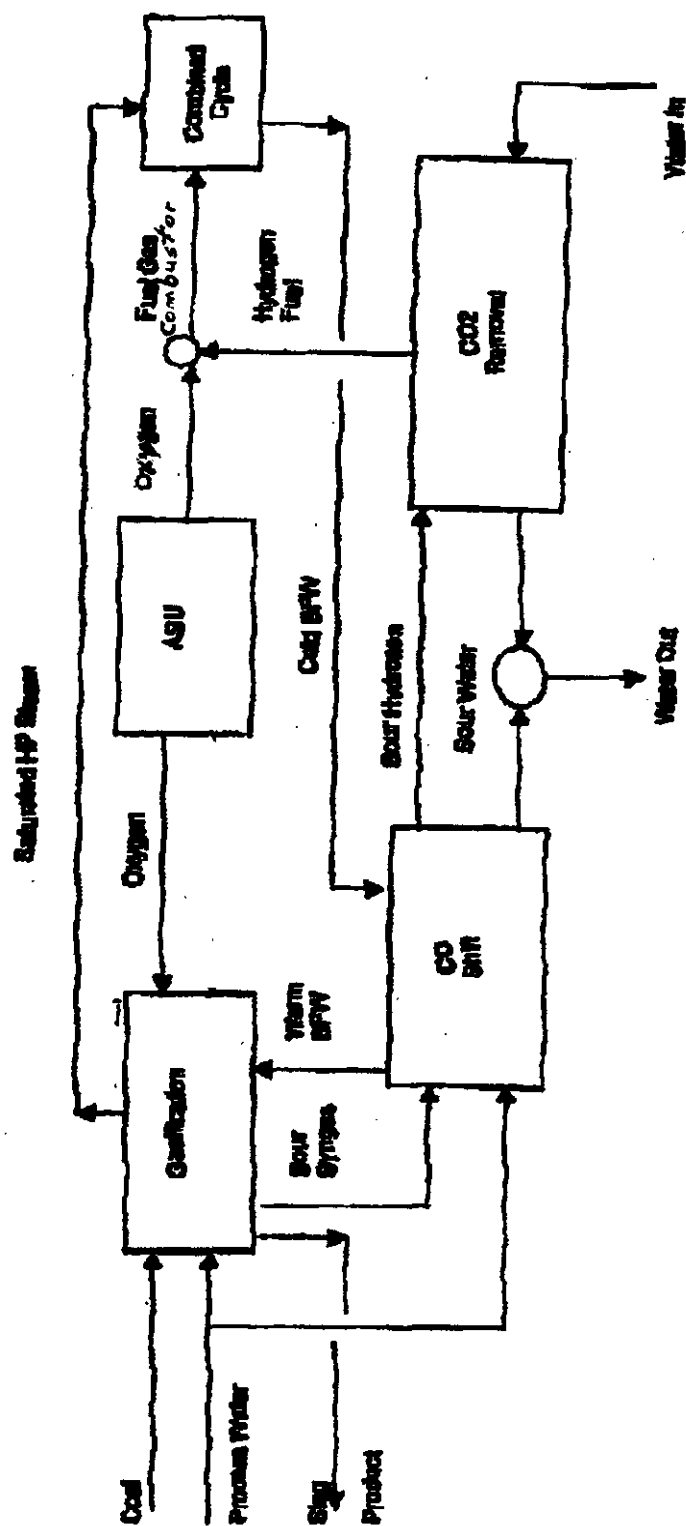
Gasification

The advanced IGCC powerplant is fed with a high sulfur bituminous coal slurry at a coal feed rate of 5000 short tons per day. Two parallel gasification trains are utilized to produce the high temperature synthesis gas (CO, H₂, CO₂ and trace gases). The gasifier and radiant syngas cooler will achieve greater than 98% carbon conversion in a single pass. Ash will be removed from the bottom of the radiant cooler utilizing Texaco's proprietary periodic flushing system.

The syngas will be futher cooled and scrubbed to remove fines from the syngas in a carbon scrubber. This "grey" water must be treated, and solids recycled to the slurry feed tanks, which provide the coal-water slurry feed for the Texaco gasifier.

FIGURE 1

DOE SMITTECHIE CO2 HYDRATE STUDY



Air Separation Unit

The Air Separation Unit (ASU) will use ambient air to produce oxygen for use both in gasification and for combustion of the hydrogen rich product stream. The ASU will be sized to produce approximately 300,000 pounds per hour of O₂ for gasification and approximately 475,000 pounds per hour for combustion. The parasitic power loads associated with air compression and refrigeration are 53 Mwe and 87 Mwe to meet these oxygen demands, respectively.

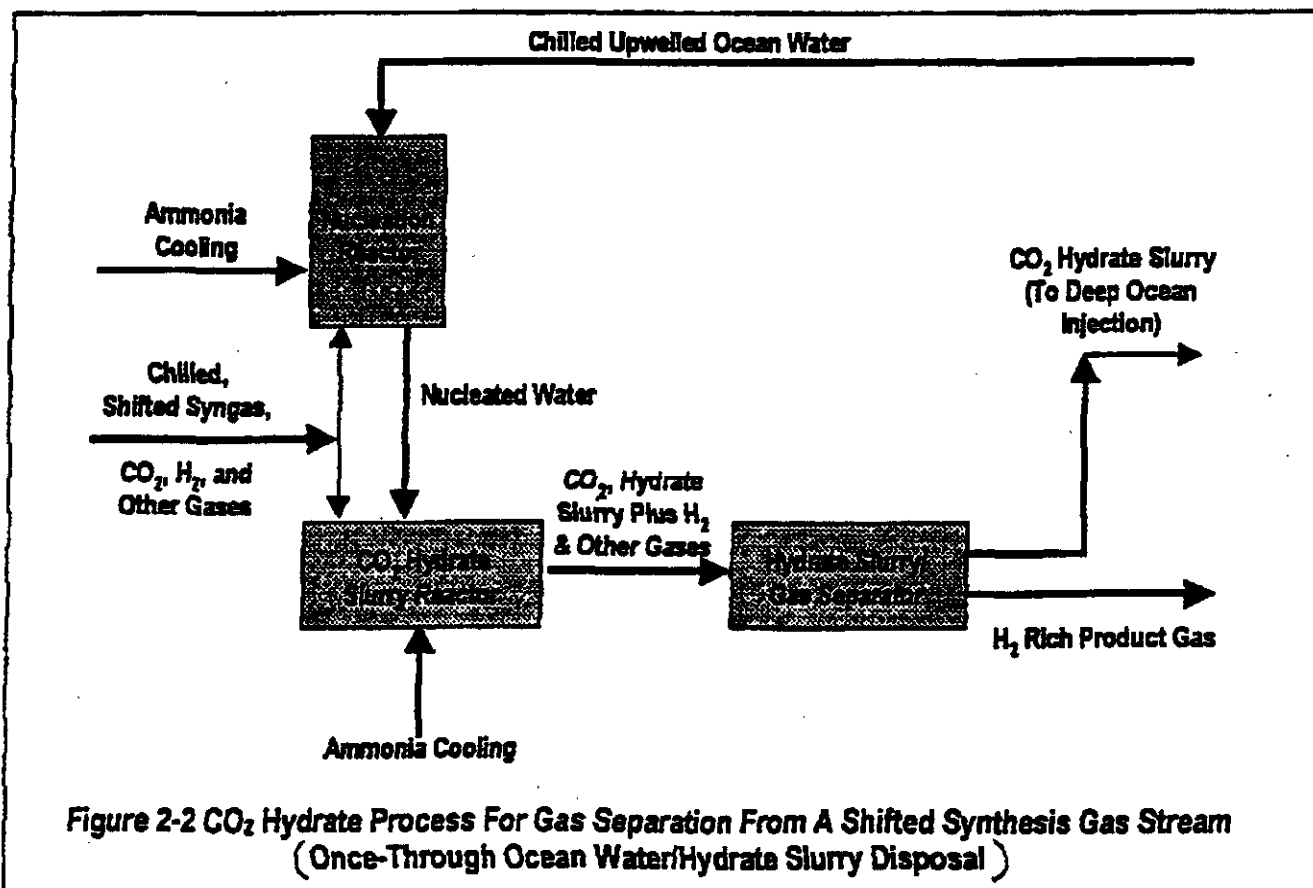
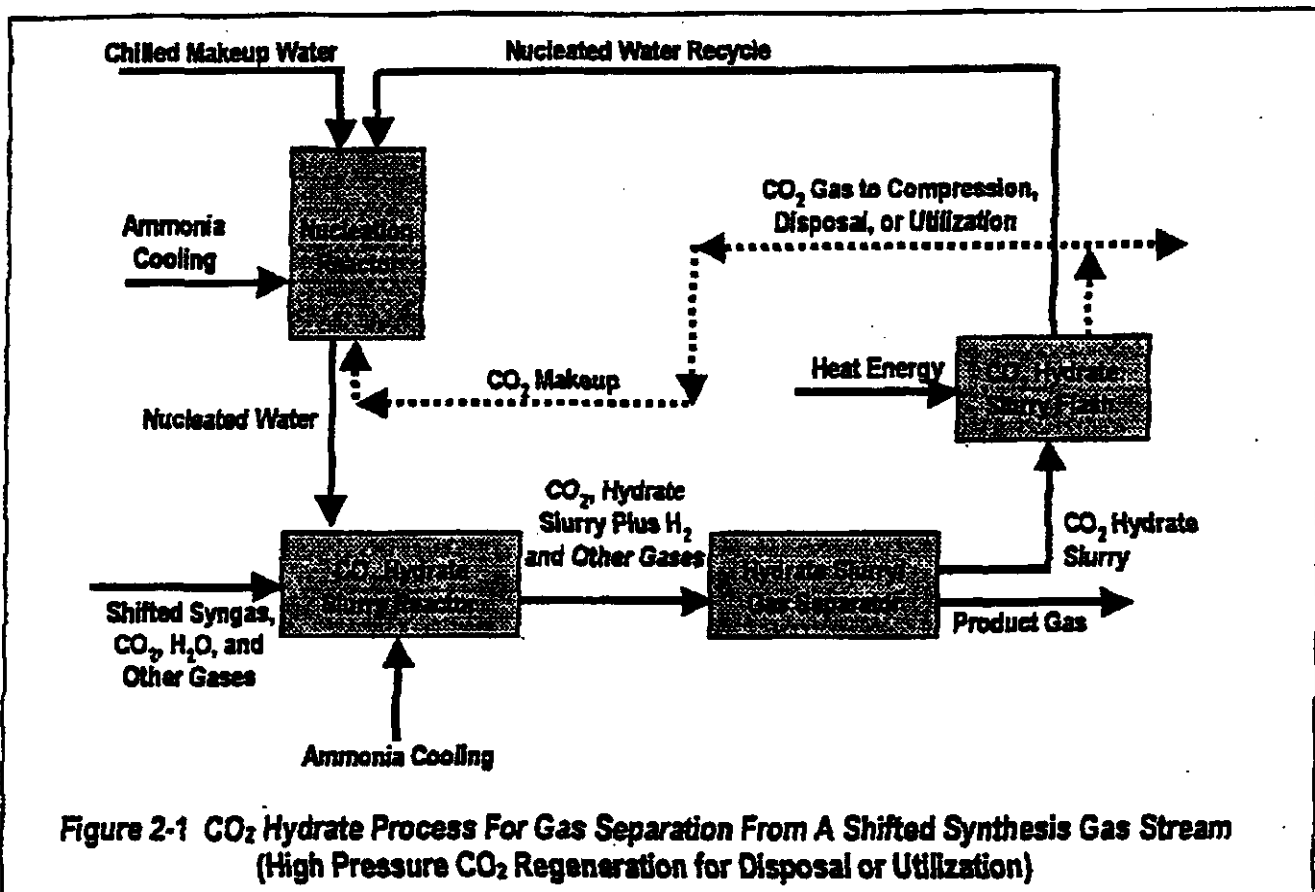
Sour Shift

In order to achieve a near 90% removal of carbon dioxide from the carbon content of the coal, it is necessary to convert the bulk of the carbon monoxide (CO) in the syngas to carbon dioxide (CO₂), and then separate nearly all of the CO₂ from the shifted synthesis gas prior to combustion. The CO is converted to CO₂ by the catalytic, water-shift reaction, wherein one mole of CO reacts with one mole of steam, over a catalyst, to produce a mole of CO₂ and a mole of H₂. This is conventional technology, even in the presence of H₂S in the syngas stream.(Ref.1) After two stages of shift conversion and cooling, and excess water is removed, the shifted syngas is further cooled to hydrate formation temperatures. The shifted synthesis gas has a composition of approximately 56 mole percent H₂, 40 mole percent CO₂, 1.7 mole percent CO, 1.1 mole percent H₂S and CO₂, and small amounts of N₂, Ar, and water vapor. The gas is now at a pressure of approximately 680 psig.

CO₂ Hydrate Separation

Figures 2-1 and 2-2 show two alternative SIMTECHE CO₂ Hydrate Processes for separation of CO₂ from a shifted synthesis gas stream. Figure 2-1 shows a process for utilization at inland plant sites, which regenerates a high pressure CO₂ stream for sequestration in ground aquifers or depleted oil and gas reservoirs or for utilization for enhanced oil recovery or coal bed methane recovery. Figure 2-2 shows the once thru upwelled ocean water hydrate separation approach which might be utilized for coastal or offshore applications, and was depicted in the overall plant process flow diagram (Figure 1).(Refs.2&3)

The present analysis focuses on the hydrate separation process shown in Figure 2-1, and includes nucleated water and partial CO₂ recycle to assure that hydrate precursors are being fed to the CO₂ hydrate slurry reactor.(Ref.4) The product gas, primarily H₂, is separated from the CO₂ hydrate slurry with over 99.6% of the hydrogen being recovered for combustion. The CO₂ hydrate is



estimated to remove approximately 85-90% of the CO₂ from the shifted synthesis gas stream, in a single pass. The remaining elements of the process are SIMTECHE Proprietary and cannot be released.

Product Gas Acid Gas Removal

The exact fate of the hydrogen sulfide (H₂S) in the shifted synthesis gas is somewhat uncertain. H₂S forms hydrates more easily than CO₂; however, based on equilibrium calculations, the partial pressure of H₂S in the shifted synthesis gas is too low for H₂S hydrates to form. Therefore, some of the H₂S will accompany the CO₂ hydrate slurry, since it will be dissolved in the slurry water and perhaps some H₂S hydrate will form. It is presently estimated that approximately 17% of the H₂S in the shifted synthesis gas, with an H₂S partial pressure of approximately 7.5 psig, will accompany the CO₂ slurry and 83% will be included in the product gas stream.

The H₂S in the product gas is only accompanied by small amounts of CO₂ i.e. approximately 2:1 on a CO₂/H₂S mole basis and, as a result, the total product gas flow rate is reduced by 40%. Therefore, the H₂S removal by an amine or Selexol solvent process is facilitated to provide clean H₂ product gas to the gas turbine combustor.

The H₂S in the CO₂ hydrate slurry is essentially all removed with the regeneration of the high pressure CO₂ stream and represents approximately 0.5 mole percent H₂S. Should H₂S hydrate form and all of the H₂S accompany the CO₂, the H₂S concentration in the CO₂ stream would be 3.0 mole percent. This is a key question which must be determined from an experimental program.

For purposes of this analysis, we assume that an additional H₂S removal process is necessary to treat the product hydrogen prior to combustion.

CO₂ Compression

The CO₂ gas is regenerated in a SIMTECHE Proprietary flash reactor. The CO₂ is regenerated at pressures of approximately 550 psig, is demineralized, and then compressed to the necessary line pressure for sequestration or utilization. Typical CO₂ sequestration/utilization pressure requirements are 1500 to 3000 psig, depending on the specific application. The CO₂ gas is also at temperatures below ambient, so compression of the gas by factors of 3 to 6 are very energy efficient. For a typical

sequestration injection pressure of 2200 psia, the CO₂ compression requirement is approximately 12 Mwe.

Overall CO₂ Separation/Regeneration/Compression Power Requirements

The power requirements for a) operating an ammonia cooling system for all gas and water cooling, including cooling of the nucleation and hydrate reactors, b) the CO₂ recycle compressor and cooler, c) the water pumping power, and d) CO₂ compression to 2200 psia have been estimated in order to evaluate the overall energy efficiency of the process. For a plant processing 5000 tons/day of coal, with 85-90% CO₂ removal, these total parasitic power requirements are less than 25 Mwe.

Power Cycle

The clean, nearly pure hydrogen (H₂) is utilized in an advanced hydrogen fueled combustion turbine cycle designed by R. Bannister of Westinghouse/Mitsubishi under contract to NEDO.(Ref.5) The so-called "Long-Term Plant Cycle" was selected, since its best integrates with the shifted synthesis gas, IGCC powerplant, and utilizes much lower high temperature combustor pressures than other proposed cycles. This long term plant cycle is estimated to be available near 2020, probably, at a time when CO₂ controls on new powerplants may be necessitated.

The Bannister Power Cycle is a four stage power train with the following key components:

- a) A supercritical steam, high pressure turbine
- b) An intermediate pressure (1015 psig) H₂-O₂ combustion/power turbine
- c) An intermediate pressure (144 psig) H₂-O₂ combustion/power turbine
- d) A supercritical steam raising Heat Recovery Steam Generator

and e) A low pressure steam turbine.

Although there are many developmental components and materials necessary for this cycle, the overall cycle efficiency is estimated to be 71.4 percent net efficiency, if the hydrogen and oxygen are provided at 1015 psig. This is the most efficient combustion turbine power cycle which I have seen utilizing hydrogen as the fuel. Although more efficient power cycles may be developed, this power cycle is the best defined, at this time.

Overall Powerplant Efficiency

Table 1 shows the estimated overall powerplant efficiency for this combined CO₂ removal/advanced IGCC power cycle plant. Even with a very high power cycle efficiency, 71.4% on a LHV basis, the overall powerplant efficiency is only 40.1% on a higher heating value basis, from coal to electricity. The gross powerplant output is estimated to be 664 Mwe, but the net powerplant output is reduced to 530 Mwe. For purposes of this analysis, we have assumed that a single power train could provide this output, but parallel trains may be necessary or desirable from a reliability/availability standpoint.

The parasitic power losses for CO₂ hydrate separation, regeneration, and compression to 2200 psig represent only 4.7% of the net output of the powerplant and are comparable to the other power requirements within the plant. The oxygen required for combustion significantly increases the air separation plant power requirements from approximately 53 Mwe for gasification (included within the cold gas efficiency of the synthesis gas) to 140 Mwe. This high power demand is required both for oxygen separation and compression to 1015 psig for combustion, and this combustion oxygen demand represent 16.4% of the net powerplant output.

Estimated Powerplant Cost and Cost of Carbon Control

The detailed plant cost analysis has not been completed for this plant design; however estimates from previous studies (Ref.6) have been utilized to provide a "first cut" cost estimate. The total estimated plant cost is \$1420 per kwe, including approximately \$100/kwe for the CO₂ separation/regeneration/compression processes, nearly \$300 per kwe for the air separation plant and oxygen compression for both gasification and combustion, and approximately \$400/kwe for the power cycle portion of the plant.

A reference IGCC plant is estimated to cost, approximately, \$1380/kwe, on the same technology basis, with a net heat rate of 7300 Btu/kwhr, compared with 8510 Btu/kwhr for the CO₂ removal case. The estimated cost of carbon control, between these two plants is \$11/ton C. If a reference pulverized coal plant is utilized, with a capital cost of \$1100/kwe and a 9000 Btu/kwhr net heat rate, the cost of carbon control is \$44/ton C. Therefore, the basic reference plant design is critical in defining the cost of carbon control. These are the in-plant costs for carbon control, but do not include pipelining and sequestration costs.

Pipelining and Sequestration Costs

Two of the sequestration options which are being considered in the U.S. are a) terrestrial aquifer storage and b) coal bed methane recovery. Recently, the literature has been reviewed

Table 1

Overall Efficiency of Advanced IGCC Powerplant
With CO₂ Removal - 5000 ST/DAY Coal

<u>Process/Cycle</u>	<u>Efficiency (HHV)</u> (Percent)	<u>Powerplant Output</u> (Mwe)
Gasifier Cold Gas Efficiency (Including O ₂ for Gasification) and Saturated Steam from Radiant Gas Cooler	76	
Catalytic Shift Reaction	96	
Powerplant Efficiency	<u>68.5</u>	
Gross Powerplant Efficiency	50.0	
Gross Powerplant Output		664
Other Parasitic Losses		(134)
* Oxygen for Combustion		(87)
* CO ₂ Separation, Regeneration & Compression		(25)
* H ₂ S Removal and Recovery		(2)
* Other Power Requirements		<u>(20)</u>
Net Powerplant Output		530
Net Powerplant Efficiency	40.1	

regarding the
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process, as

Terrestrial

An IEA Green
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the methane, including CO₂ compression costs.

Table 2 summarizes the key properties of two alternative terrestrial disposal/sequestration or utilization options. Clearly CBM recovery is a preferred approach for regions where this option is available (Southwestern and Southeastern U.S. primarily). Terrestrial aquifers, which are ubiquitous to most coal producing regions provide a much larger reservoir potential, but at potentially high cost, up to \$67 per ton of C. If the lower end costs can be achieved, this may also be a viable option.

Conclusions

Table 3 shows a total cost of carbon control based on the range of cost assumptions, both within the plant for CO₂ hydrate separation, regeneration and compression and for the two sequestration options discussed above. The key differences in the carbon control costs are:

1. The reference plant basis utilized to estimate the difference from the CO₂ control plant with CO₂ hydrate separation, regeneration and compression costs versus the reference powerplant, with no CO₂ control. If the reference design is another IGCC plant, the cost of carbon control is estimated to be approximately \$11/ton C, within the plant gates.

The two alternative sequestration/utilization approaches produce net costs of carbon control from \$6-11 per ton C, utilizing coal bed methane and the IGCC reference plant, to \$49 to \$111 per ton C utilizing terrestrial aquifer sequestration and a reference supercritical coal fired powerplant. Therefore, in quoting carbon control costs from advanced coal power stations, it is extremely important to clearly identify the reference plant bases which are being utilized for comparison purposes.

Table 2

Comparison of Two Potential Terrestrial CO₂ Disposal/
Sequestration or Utilization Approaches in the U.S.

<u>Property</u>	<u>Terrestrial Aquifers</u>	<u>Coal Bed Methane Recovery</u>
Acceptable H ₂ S Concentration, %	100%/0% to 32%/68%	100%/0% to 97%/3%
Injection Pressure, psig	1400-3000	1450-2200
Injection Temperature, °F	70-110	70-110
U.S. CO ₂ Storage Potential	100 GT CO ₂	30 GT CO ₂
Estimated Cost for CO ₂ Sequestration, \$/Ton C	5-67	-5 to 0

Table 3

Estimated Carbon Control Costs for Nominal 500 Mwe
Shifted Synthesis Gas IGCC Plant - Today's Dollars

<u>Function</u>	<u>Cost Per Ton of Carbon, \$/Ton C</u>	
	<u>IGCC Reference Plant</u>	<u>Advanced Pulverized Coal Plant</u>
Inplant CO ₂ Hydrate Separation, Regeneration, and Compression	11	44
Terrestrial Aquifer Sequestration	<u>5-67</u>	<u>5-67</u>
Total Carbon Control Cost	16-78	49-111
In Plant CO ₂ Hydrate Separation, Regeneration, and Compression	11	44
Coal Bed Methane Recovery	<u>-5-0</u>	<u>-5-0</u>
Total Carbon Control Cost	6-11	39-44

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